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Dept: Civil Engineering
Thesis Title: Feasibility Study of Upflow Anaerobic Filter for Pretreatment of Municipal Wastewater

SUMMARY

Anaerobic reactors have been successfully installed in full-scale plants world-wide for treating high-strength industrial wastewater over the years. Recently, there has been significant interest in exploring this technology for treating low-strength domestic wastewater as well. Previously, it was thought that this was not practical as methane fermentative process was considered too slow to be able to treat the increasing volume of domestic sewage at a high rate. With technological advances and better understanding of anaerobic microbial characteristics in recent years, there is a potential that under control conditions, such barriers can be gradually overcome. The perspectives of using anaerobic pre-treatment for domestic sewage are discussed in this report to replace the conventional treatment methods. Feasibility of upflow anaerobic filter (UAF) in place of activated sludge process to pre-treat domestic wastewater is studied in this research.

Keywords: Anaerobic Filter, Sewage, COD, BOD, TSS and Methane.
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The last but not least, I am indebted to my husband, parents, sisters, brothers and my son for their care, support and sacrifices to finish my research successfully.
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<td>Activated sludge process</td>
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<tr>
<td>BOD</td>
<td>Biochemical Oxygen Demand</td>
</tr>
<tr>
<td>COD</td>
<td>Chemical Oxygen Demand</td>
</tr>
<tr>
<td>HRT</td>
<td>Hydraulic Retention Time</td>
</tr>
<tr>
<td>sBOD</td>
<td>Soluble BOD</td>
</tr>
<tr>
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<td>VFA</td>
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<td>Volatile Suspended Solids</td>
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CHAPTER 1  INTRODUCTION

1.1 Background

With increasing world population and demand for more fresh water, recycling and reuse of wastewater has gained popularity among countries and industries in recent years. Water reclamation on treated effluent has now been considered as one of the alternative sources for water, especially for regions that face water scarcity issues. Therefore, reclamation of wastewater has the double advantage of reducing the demand for fresh water and protecting the water quality of the receiving bodies.

About one-third of the world's population lives in countries with moderate to high water stress, and problems of water scarcity are increasing, partly due to ecosystem depletion and contamination. Two out of every three persons on the globe may be living in water-stressed conditions by the year 2025, if present global consumption patterns continue (WHO, 2000). Meanwhile, water consumption has increased nine fold and industrial water consumption has risen by a factor of 40. Yet water as a resource is limited and poorly distributed. "The quantity of available water remains the same. Its scarcity could be a serious obstacle to development in the millennium” (GEO, 1999).

For decades, Singapore has relied on import from Malaysia to supply half of the water consumption in Singapore. However the two water agreements that supply Singapore this water are due to expire by 2011 and 2061, respectively, and the two countries are engaged in an on-going discussion over the price of raw water. Without a workable resolution, the government of Singapore decided to increase self-sufficiency
in its water supply. Once wastewaters are produced and collected in sewerage systems, treatment becomes a necessity. Yet, wastewater management is a costly business. Water reclamation plants in Singapore treated about 511 million cubic meters of used water in the year 2006 (PUB, 2006). The Keppel Seghers Ulu Pandan NEWater Plant has a capacity to produce 32 mgd (148,000 m$^3$/day) of NEWater to supply over 50 percent of Singapore’s current NEWater needs. Therefore, water reuse can be a better option to solve water requirement in tropical countries like Singapore to some extent.

The current wastewater treatment in Singapore follows mainly the conventional treatment train. Most conventional wastewater treatment processes are aerobic; that is, the bacteria used to break down the waste products take in oxygen to perform their function. This results in high energy consumption, huge land area requirement and a large volume of waste sludge being produced. Indeed, treatment and disposal of sewage sludge is technically cumbersome and economically a heavy burden. This makes the processes complicated to control and costly to operate. To overcome these problems, anaerobic treatment system can be an alternative to treat domestic wastewater in Singapore, where land and sludge disposal is a major concern.

The bacteria in anaerobic processes do not use oxygen. Therefore, the energy requirement and sludge production are much lesser than aerobic processes, making anaerobic processes becoming cheaper alternative. Ng and Chin (1987) reported that anaerobic digestion processes are energy efficient as they do not need to transfer large quantities of oxygen into the wastewater. Sludge management requirements are also
reduced because the process produces substantially less biological solids than conventional aerobic treatment processes. In addition, the methane-rich biogas generated by the process is a convenient energy source for plant operation.

It is often questioned why aerobic treatment of municipal wastewater is not replaced more rapidly by the economically more attractive and the conceptually more holistic anaerobic treatment. Also, the temperature range conducive for bacteria is very much suited for hot climates like in Singapore.

Anaerobic reactors have been successfully installed in full-scale plants worldwide for treating high-strength industrial wastewater over the years. Recently, there has been significant interest in exploring this technology for treating low-strength domestic wastewater as well. Previously, it was thought that anaerobic process was not practical as methane fermentative process was considered too slow for treating the increasing volume of domestic sewage at a high rate. With technological advances and better understanding of anaerobic microbial characteristics in recent years, there is a potential that under control conditions, such barriers can be gradually overcome. The perspectives of using anaerobic pre-treatment for domestic sewage are discussed in this report to replace the conventional treatment methods.
1.2 Objectives and Scope of the Study

The scopes of this study covered:

1. Feasibility study on using selected anaerobic treatment technology in place of activated sludge process to pre-treat domestic wastewater using bench-scale systems.

2. Performance examination of an upflow anaerobic filter (UAF) at hydraulic retention times of 16, 12, 8, 6 and 4hrs.

3. Optimize the anaerobic processes for maximal energy production and organic removal.

The specific objectives of the study were: (1) to determine the stability of the process at short HRTs, (2) to examine its treatment efficiencies, and (3) to compare process parameters and performances with other studies.
CHAPTER 2 LITERATURE REVIEW

2.1 Anaerobic Treatment Technology

2.1.1 Fundamentals of Anaerobic Decomposition

2.1.1.1 Anaerobic Bacteria

Anaerobes (literally meaning "without air") are organisms that do not use oxygen to live. Anaerobic organisms use different molecules as electron acceptors, such as sulfide or carbon dioxide. In fact, these organisms are incredibly diverse when it comes to the nutrients that they can use to survive.

2.1.1.2 Pathways in Anaerobic Degradation of Organic Waste

The use of anaerobes in the absence of oxygen for the stabilization of organic material by conversion to methane, carbon dioxide, new biomass and inorganic products is known as anaerobic degradation. There are three distinct phases, namely, hydrolysis, acidogenesis and methanogenesis. The diagram of the process of anaerobic degradation is presented in Figure 2.1.

The anaerobic process is different from the aerobic process in a way that it occurs in the absence or very low amounts of oxygen such that aerobic reactions, in which oxygen act as the electron acceptors, cannot take place. This process involves 4 main phases where different types of bacteria, which will be mentioned below, convert large complex organics into smaller compounds such as methane. These bacteria depend on each other to achieve a balanced growth. The breakdown of organics under anaerobic condition is given in Equation 2.1.
Anaerobic overall equation:

\[ \text{Organics} \rightarrow \text{CH}_4 + \text{CO}_2 + \text{H}_2 + \text{NH}_3 + \text{H}_2\text{S} \quad (2.1) \]

Figure 2.1 Pathways in Anaerobic Degradation

**Hydrolysis**

Hydrolysis is the first step in the anaerobic process, in which particulate matter is converted to soluble compounds that can be hydrolyzed further to simple monomers that are used by bacteria that perform fermentation. This step is necessary to allow the organic materials to pass through the bacterial cell walls for use as energy to meet metabolic requirements. This is done by the excrement of extra-cellular and hydrolytic enzymes. In the anaerobic processes, hydrolysis would be best described as a first order process with respect to the concentration of degradable particulate organic matter. Table 2.1 shows the list of hydrolytic bacteria and extra-cellular enzymes that involved in hydrolysis process.
to degrade complex organic compounds such as protein, carbohydrate and lipid (Maier et al., 2000).

Factors which affect the rate of hydrolysis include pH, sludge retention time (SRT) and particulate size of substrate. In hydrolysis, there is no reduction of chemical oxygen demand (COD) as macromolecules are merely broken down into monomers. Soluble COD is expected to increase due to the hydrolysis of macromolecules into soluble organic products.

Table 2.1 List of hydrolytic bacteria and extracellular enzymes (Maier et al., 2000)

<table>
<thead>
<tr>
<th>Complex Organic Compound</th>
<th>Hydrolytic Bacteria</th>
<th>Extracellular Enzyme</th>
</tr>
</thead>
<tbody>
<tr>
<td>Protein</td>
<td><em>Clostridium, Bacillus</em>, <em>Vibrio Peptococcus</em></td>
<td>Protese</td>
</tr>
<tr>
<td>Carbohydrate</td>
<td><em>Clostridium, Acetovibrio celluliticus</em>, <em>Bacteridies</em></td>
<td>Cellulase, Amylase, Xylanase</td>
</tr>
<tr>
<td>Lipid</td>
<td><em>Clostridium, Micrococcus</em></td>
<td>Lipase, Phospholipase</td>
</tr>
</tbody>
</table>

**Acidogenesis**

Acidogenesis is the second step in the anaerobic process. The complex organic matter that has been hydrolyzed ferment to long chain organic acids, sugars and amino acids, after which, they are degraded further. Organic substances serve the function of both electron acceptors and donators. The principal products are acetate, hydrogen, carbon dioxide, propionate and butyrate. This stage of the anaerobic degradation is mediated by facultative and obligate bacteria. Studies have shown that the obligate anaerobes form the
larger portion of the acidogenic bacteria as compared to the facultative anaerobes (Maier et al., 2000).

In this step, COD reduction is due to the conversion of soluble organics to biomass and to biogas in the form of carbon-di-oxide (CO$_2$) and hydrogen (H$_2$). Most of the COD is still in the soluble state; however, it has been changed to acetate and other volatile fatty acids (VFAs). Table 2.2 shows the list of acidogenic bacteria involved during acidogenesis process (Maier et al., 2000).

Table 2.2 List of acidogens involved in acidogenesis (Maier et al., 2000)

<table>
<thead>
<tr>
<th>Source Product</th>
<th>Acidogen</th>
<th>Product</th>
</tr>
</thead>
<tbody>
<tr>
<td>Long chain fatty acids, glycerol</td>
<td>Clostridium</td>
<td>Higher VFAs, Acetate, H$_2$, CO$_2$</td>
</tr>
<tr>
<td>Sugar, amino acids</td>
<td>Clostridium, Streptococcus, Eubactrium limosum</td>
<td>Higher VFAs</td>
</tr>
<tr>
<td></td>
<td>Zymomonas mobilis</td>
<td>Long chain fatty acids, alcohol</td>
</tr>
<tr>
<td></td>
<td>Lactobacillus, Micrococcus, Escherichia, Pseudomonas, Staphylococcus, Bacillus, Desulfovibrio, Selenomonas, Veillonella, Sarcina, Streptococcus, Desulfobacter, Desulfuromonas</td>
<td>Acetate, H$_2$, CO$_2$</td>
</tr>
</tbody>
</table>
Acetogenesis

This stage is a subset of acidogenesis and involves the oxidation of long chain fatty acids, propionate and butyrate by obligate anaerobes to produce hydrogen, carbon dioxide and acetate. Even number carbon atom acids are degraded to acetate, whereas odd number carbon acids are degraded to acetate and hydrogen ion (H\(^+\)).

Thus the final products of acidogenesis are the precursors of methane production. In this stage, there is little or no stabilization but only a change in the form of the organic material. Table 2.3 shows the list of acetogenic bacteria involved in acetogenesis (Maier et al., 2000).

<table>
<thead>
<tr>
<th>Source</th>
<th>Acetogens</th>
<th>Product</th>
</tr>
</thead>
<tbody>
<tr>
<td>Long chain fatty acids, alcohol</td>
<td>Syntrophomonas wolfei</td>
<td>Acetate, H(_2), CO(_2)</td>
</tr>
<tr>
<td>Higher VFAs</td>
<td>Syntrophomonas wolfei, Syntrophomonas wolinnii</td>
<td>Acetate, H(_2), CO(_2)</td>
</tr>
</tbody>
</table>

Methanogenesis

Methanogenesis is the third step in the anaerobic process. There are two main groups of methanogens that are responsible for this, namely aceticlastic methanogens and hydrogen utilizing methanogens. The aceticlastic methanogens are responsible for splitting the acetate into methane and carbon dioxide (Metcalf and Eddy, 2003). Eq. (2.1) shows the
splitting of acetate into methane and carbon dioxide while Eq. (2.1) shows the reduction of carbon dioxide in the presence of hydrogen.

Acetotrophic:

\[ \text{CH}_3\text{COOH} \rightarrow \text{CH}_4 + \text{CO}_2 \quad \Delta G^\circ = -32 \text{ kJ} \quad (2.2) \]

Hydrogenotrophic:

\[ \text{CO}_2 + 4\text{H}_2 \rightarrow \text{CH}_4 + 2\text{H}_2\text{O} \quad \Delta G^\circ = -138.9 \text{ kJ} \quad (2.3) \]

The second group use hydrogen as the electron donor and carbon dioxide as the electron acceptor to produce methane. These are the acetogens which are also able to use carbon dioxide to oxidize hydrogen and form acetic acid. Methane fermentation is very important in the anaerobic treatment process. Stabilization of the organic material occurs when acetic acid is converted to methane. In general, about 72% of methane produced in an anaerobic process is from acetate formation (Metcalf and Eddy, 2003). The other 28% is contributed by the reduction of carbon dioxide using hydrogen as the energy source by carbon dioxide reducing bacteria (Henze et al., 1983; Parkin et al., 1986). It is also noted that high concentrations of propionate or butyrate is indicative of reactor failure, and propionate, in particular, is toxic to the acetogens (Parkin and Owen, 1986). Table 2.4 shows the list of methanogens involved in methanogenesis (Maier et al., 2000).

<table>
<thead>
<tr>
<th>Source Product</th>
<th>Methanogen</th>
<th>Product</th>
</tr>
</thead>
<tbody>
<tr>
<td>Acetate</td>
<td><em>Methanotrix</em>, <em>Methanosarcina</em>, <em>Methanospirillum</em></td>
<td>CH(_4), CO(_2)</td>
</tr>
</tbody>
</table>

Table 2.4 List of methanogens involved in methanogenesis (Maier et al., 2000)
2.1.2 Kinetics of Anaerobic Decomposition

Process kinetics has been used for the mathematical description of both aerobic and anaerobic biological treatment processes. The understanding of process kinetics is essential for the rational design and operation of any biological waste treatment and for predicting system stability, waste stabilization and effluent quality.

Many attempts have been made to formulate expressions to describe the kinetics of micro-organism metabolism. Many of these expressions are based on work carried out by Monod (Metcalf and Eddy, 2003), who studied the fermentation of grape sugars to alcohol. The results of the work of Monod can be summarized by two basic principles:

1. the growth rate of the micro-organisms, which was found to be proportional to the rate of substrate utilization:

\[
(dX/dt)_g = Y(dS/dt)_u = X_{\mu} = X_{\mu_m}S/(S+K_s) \tag{2.4}
\]

2. the decay rate of the micro-organisms, which can be expressed by a first order equation:

\[
(dX/dt)_d = -X_b \tag{2.5}
\]

where \(X\) = microorganism concentration (mg VSS/L); \(S\) = substrate concentration (mg COD/L); \(\mu\) = specific growth rate of microorganisms (1/d); \(\mu_m\) = maximum specific growth rate (1/d); \(b\) = death rate constant (1/d); \(K_s\) = Monod constant (mg COD/L). From equation (2.4) it follows that, at high substrate concentrations, the Monod ratio \(S/(S+K_s)\)
approaches unity and the growth rate becomes independent of the substrate concentration, i.e. it becomes a zero-order process. If the substrate concentration is low ($S \ll K_s$), the Monod ratio approaches $S/K_s$ and the growth rate is proportional to the substrate concentration, which is characteristic of a first-order process. For intermediate concentrations the growth rate is between zero and first order with respect to the substrate concentration.

The specific growth rates of *Methanotrix* and *Methanosarcina* are 0.1 and 0.3 d$^{-1}$, respectively (Adrianus and Lettinga, 1994). The specific growth rate is at half its maximum value when the substrate concentration is equal to the parameter $K_s$, which, for that reason, is called the half-saturation constant or affinity constant. For *Methanotrix* and *Methanosarcina* the values of $K_s$ are 200 and 30 mg/L acetate, respectively. At low acetate concentration ($<55$ mg/L) the specific growth rate of *Methanotrix* becomes higher than that of *Methanosarcina*. By contrast, at acetate concentrations exceeding 55 mg/L, *Methanosarcina* will out-compete *Methanotrix* and become the prevailing acetate-consuming organism.

In sewage treatment practice the substrate concentration will not be the minimum obtainable, because this would require a very long retention time and hence an unacceptably large treatment process. If the substrate concentration is greater than the minimum there will be a net growth of microorganisms. Naturally, the increase in the microorganism mass cannot go on indefinitely: after some time of operation the system will be full and wastage of microorganism mass becomes unavoidable. If it is assume that
the microorganisms produced in a completely mixed treatment system are wasted at a
constant rate, this rate will be equal to the net production rate. In that case a constant
microorganism mass and concentration, compatible with the organic load entering the
system, will establish itself. The rate of wastage is the inverse of the sludge age, which
denotes the average solids retention time. Thus for a steady-state system

\[
(dX/dt)_w = (dX/dt)_g + (dX/dt)_d
\]  
(2.6)

Or \[ X/R_s = X(\mu - b) \]  
(2.7)

where \( X \) = microorganism concentration (mg VSS/L)

\( (dX/dt)_w \) = rate of wastage

\( (dX/dt)_g \) = growth rate of the micro-organisms

\( dX/dt)_d \) = decay rate of the micro-organisms

\( R_s \) = Sludge age

The following expression is for the effluent substrate concentration:

\[
S = K_s(b+1)/R_s/[(\mu_m - (b+1/R_s)]
\]  
(2.8)

Equation (2.8) shows that the effluent concentration depends upon the values of three
constants (\( K_s, \mu_m \) and \( b \)) and one process variable: sludge age, \( R_s \).

Another important kinetic parameter is the maximum specific substrate utilization rate,
\( K_m \). This constant denotes the maximum mass of substrate that can be metabolized per
unit time. Specific substrate utilization rate can be calculated from the maximum specific
growth rate and the yield coefficient as follows:

\[
K_m = \mu_m / Y
\]  
(2.9)

\( K_m \) = specific substrate utilization rate (kg COD/kg VSS/d)
Henze and Harremoes (1983) estimated the most important kinetic constants for acid and methanogenic fermentation from the results of a large number of experimental investigations. The values are presented in Table 2.5.

<table>
<thead>
<tr>
<th>Cultures</th>
<th>( \mu_m ) (d(^{-1}))</th>
<th>Y (mg VSS/mg COD)</th>
<th>( K_m ) (mg COD/mg VSS.d)</th>
<th>( K_s ) (mg COD/L)</th>
</tr>
</thead>
<tbody>
<tr>
<td>Acid-producing bacteria</td>
<td>2.0</td>
<td>0.15</td>
<td>13</td>
<td>200</td>
</tr>
<tr>
<td>Methane-producing bacteria</td>
<td>0.4</td>
<td>0.03</td>
<td>13</td>
<td>50</td>
</tr>
<tr>
<td>Combined culture</td>
<td>0.4</td>
<td>0.18</td>
<td>2</td>
<td>-</td>
</tr>
</tbody>
</table>

In principle, it is an advantage to increase the sludge age by retaining the sludge in the reactor system. There is, of course, a practical limit, because there will be maximum sludge concentration in the treatment system, so the sludge can only be retained if the reactor volume is sufficiently large. It is concluded that a treatment system can only be efficient if a large sludge concentration can be maintained in it.

### 2.1.3 Factors Affecting Anaerobic Treatment

#### Temperature

Anaerobic digestion, like other biological processes, strongly depends on temperature. Microorganisms are classified into temperature classes on the basis of the optimum
temperature and the temperature span in which the species are able to grow and metabolize. Figure 2.2 shows the various methonogens and their growth rates.

![Figure 2.2 Growth rates of methanogens (Lettinga et al., 2001).](image)

A strong temperature effect on the maximum substrate utilization rates of microorganisms has been observed by many researchers (Lettinga et al., 2001). In general, lowering the operating temperature leads to a decrease in the maximum specific growth and substrate utilization rates but it might also lead to an increased net biomass yield (g biomass/g substrate converted) of methanogenic population or acidogenic sludge (Lettinga et al., 2001). A drop in temperature is accompanied with a change of the physical and chemical properties of the wastewater, which can considerably affect design and operation of the treatment system. For instance, the solubility of gaseous compounds increases as the temperature decreases below 20°C. At low temperatures, the liquids viscosity is also increased. Therefore, more energy is required for mixing and sludge bed reactors become less easily mixed, particularly at low biogas production rates. Henze and Harremoes (1983) concluded that the optimum temperature range is between 30 and 40°C.
and for temperatures below the optimum range the digestion rate decreases by about 11% for each degree of temperature decrease, or according to the Arrhenius expression as shown in equation 2.10

\[ r_t = r_{30}(1.11)^{(t-30)} \]  

(2.10)

where \( t = \) temperature in °C and \( r_t, r_{30} = \) digestion rate at temperature \( t \) and 30°C, respectively. The influence of temperature on anaerobic digestion is not limited by the rate of the process; the extent of anaerobic digestion is also affected.

**pH**

The value and stability of the pH in an anaerobic reactor is extremely important because methanogens can be grown at near neutral pH conditions (6.5-8.2), (Adrrianus and Lettinga, 1994; Buyukkamaci *et al.*, 2004). At pH values below 6.3 or above 7.8, the rate of methanogenesis decreases. Acidogenic populations are significantly less sensitive to low or high pH values and hence acid fermentation will prevail over methanogenic fermentation, which may result in souring of the reactor contents.
2.1.4 Advantages of Anaerobic Treatment Systems

Figure 2.3 shows the advantages of anaerobic treatment in comparison with the aerobic treatment system.

The draw towards anaerobic treatment systems over aerobic treatment systems for treating domestic sewage are summarized as follows (Ng and Chin, 1987; Mergaert et al., 1992; Van Haandal and Lettinga, 1994; Bodik et al., 2000; Mohammad and Vinod, 2000; Bodik et al., 2002; Pravin et al., 2002; Bodik et al., 2003; Mahmoud et al., 2003; Metcalf & Eddy, 2003; Omil et al., 2003; Chernicharo and Sperling, 2005).

1. Low production of excess sludge
2. Low nutrient requirements
3. No energy requirements for aeration

5. The process can accept high organic loading rates (OLRs) since oxygen transfer is not a limiting factor as aerobic process.

6. Anaerobic sludge can be preserved, unfed for many months without any serious deterioration.

7. Valuable compounds like ammonia are conserved, which in specific cases might represent an important benefit i.e. if irrigation can be applied.

2.2 Positive Perspectives for Applicability of Anaerobic Sewage Treatment

Anaerobic treatment has also found widespread application for various industrial wastewaters, like sugar beet, slaughterhouse, starch, brewery wastewaters, etc. The supporting factors of sewage for the applicability of anaerobic processes are described in the following sections.

2.2.1 Temperature in Tropical Countries

The applicability of anaerobic treatment for domestic sewage depends strongly on the temperature of sewage. The activity of mesophilic anaerobic bacteria is at its optimum at 35°C (Van Haandel and Lettinga, 1994). At lower temperatures, bacterial activity decreases, which results in lower treatment performances. This is the reason why in cold climate countries, only a small separated portion of the sewage, namely the primary and secondary sludge are treated anaerobically, however requiring heavy insulation and heating system, while the bulk of the volume, the wastewater, is treated aerobically.
mostly with aerators in open and closed ponds (Van Haandel and Lettinga, 1994). Figure 2.4 illustrates the critical temperature ranges grey shaded areas indicating sewage temperatures of 12 - 15°C, the areas between the dotted lines temperature above 20°C.

![Figure 2.4 World temperature zones (Van Haandel and Lettinga, 1994)](image)

Consequently anaerobic sewage treatment is primarily of interest for countries with a tropical or sub-tropical climate, which are mostly developing countries. Bodik et al., (2000) studied a lab-scale upflow anaerobic filter and pilot-scale anaerobic baffled filter to treat municipal wastewater and they found that:

1) Anaerobic wastewater treatment process is suitable for municipal or domestic wastewater.

2) COD removal efficiency was dependant mainly on temperature and HRT. Under low values of HRT, the removal efficiency was significantly influenced by temperature.
3) The lab-scale model was operated without any technological problem. The start-up process was realized at 23°C and was very rapid (i.e., two weeks).

4) Under ambient temperature, it was possible to obtain relatively high COD and 5 day biochemical oxygen demand (BOD$_5$) removal efficiency.

5) Decrease in COD and BOD$_5$ removal efficiencies were observed with decreasing temperature.

2.2.2 Wastewater Organic Strength

Speaking of this technology, in addition to appropriate sewage temperatures, a further precondition for effective anaerobic treatment is the organic strength of the wastewater. The initial organic strength should be above 250 mg COD$_{in}$/l, the optimum strength being > 400 mg COD$_{in}$/l (Technical Information W3e, 2001). Derin et al., (1997) mentioned that the BOD$_5$: COD ratio, conventionally regarded as an index of biological treatability is calculated as 0.47. And similarly, for the COD:N ratio, a parameter closely related to the denitrification potential, experimentally results converged to a mean value of 9.2, practically the same as the limit below which pre-denitrification is favoured.

The low organic strength in domestic wastewaters (250 – 1000 mg COD/L) has to be considered relative to the high threshold value of the methane producing bacteria. The work done by Fukuzaki et al., (1990) shows that methanogens experienced a lower substrate limit which they do not function properly. This so-called threshold related to undissociated acetic acid, the true substrate for acetogenic methanogens. This could easily result in residual volatile fatty acids (VFA) levels which are high with respect to the levels of the incoming sewage and thus implicate low removal efficiency.
Consequently, anaerobic treatment is of interest only for relatively concentrated domestic wastewaters (COD ≥ 500 mg/L) unless in the case of highly adapted *Methanotrix* sludges. Sewage characteristics which can have direct implications on the anaerobic process are summarized in Table 2.6.

Table 2.6 Composition ranges of municipal wastewater for industrialized countries (Mergaert *et al*., 1992 and Metcalf & Eddy, 2003)

<table>
<thead>
<tr>
<th>Characteristics</th>
<th>Average</th>
</tr>
</thead>
<tbody>
<tr>
<td>Total Chemical Oxygen Demand (tCOD), mg/L</td>
<td>500</td>
</tr>
<tr>
<td>Total Kjeldahl Nitrogen (TKN), mg/L</td>
<td>50</td>
</tr>
<tr>
<td>Ammonical Nitrogen (NH$_4^+$-N), mg/L</td>
<td>25-40</td>
</tr>
<tr>
<td>Volatile acids as acetic acid, mg/L</td>
<td>40</td>
</tr>
<tr>
<td>Sulphate, (SO$_4^{2-}$), mg/L</td>
<td>75</td>
</tr>
<tr>
<td>Lipids, mg/L</td>
<td>40-100</td>
</tr>
<tr>
<td></td>
<td>(Alves <em>et al</em>., 2001)</td>
</tr>
</tbody>
</table>

### 2.2.3 Total Kjeldahl Nitrogen (TKN) and Ammonical Nitrogen (NH$_4^+$-N)

The NH$_4^+$ concentration in domestic wastewater is in the range of 25 – 40 mg/L (Mergaert *et al*., 1992). This represents no problem for anaerobic treatment. The ratio of COD: N of 100:10 for domestic wastewater, is also higher than the minimum amount of nitrogen necessary for normal anaerobic sludge growth (ratio COD: N = 100:1.25) (Mergaert *et al*., 1992). The COD: N: P ratio of 100:13:2 indicated the high treatability of the wastewater by an anaerobic process. Panswad and Komolmethee (1997) indicated...
that the optimum nutrient ratio given as COD: N: P was 190 to 350:5:1. Anaerobic treatment however being feasible up to a ratio of 100:5:1. This shows that the average sewage composition meets these requirements.

### 2.2.4 Fatty Acids

The relatively low levels of VFA coupled to the alkalinity of domestic wastewater make it unlikely that inhibition by VFA has to be a concern. Long chain fatty acids, e.g. from soaps, appear to be more toxic (50% inhibition at 500 mg/L; Mergeart et al., 1992) and can sometimes be present in domestic waste as a result of certain seasonal household habits. This aspect necessitates further research.

### 2.2.5 pH

Kobayashi et al. (1983) studied a laboratory scale anaerobic filter for treatment of low strength domestic wastewater which had pH in the range of 5.72 to 8.95 with an average of 7.51. In addition to that it was reported that the pH of treated domestic wastewater was in range of 6.85 to 8.2 with an average of 7.28. Methanogens can be grown at near neutral pH conditions, defined as 6.5 - 8.2, which is a normal pH value of sewage (Buyukkamaci et al., 2004). The average sewage composition meets these requirements.

### 2.2.6 Sulfate

The sulfate levels in domestic wastewater are relatively low so it is unlikely that the critical value of 50 mg/L hydrogen sulphide (H₂S). Since the optimal temperature for
sulfate reducing bacteria (SRB) is between 30 and 35°C and thus a little lower than the optimal temperature for methane producing bacteria (MPB) (between 35 and 45°C) (Mergeart et al., 1992), it is possible that at sewage temperatures of 10 - 20°C the SRB tend to out compete the MPB so that a major part of the COD is consumed for sulfate reduction with contaminant production of corrosive sulfides. Hence, direct anaerobic treatment of municipal wastewater will necessitate post-treatment. MetCalf and Eddy (2003) indicated that the concentration of oxidized sulfur compounds in the influent wastewater to an anaerobic treatment process is important, as high concentrations can have a negative effect on anaerobic treatment. As mentioned earlier sulfate reducing bacteria compete with the methanogenic bacteria for COD and thus can decrease the amount of methane gas production. While low concentrations of sulfide (less than 20 mg/L) are needed for optimal methanogenic activity, higher concentrations can be toxic. Methanogenic activity was reported to decrease by 50% or more at H₂S concentrations ranging from 50 to 250 mg/L (Mergeart et al., 1992).

2.2.7 Toxicants

Control of toxicants is also an important issue in the anaerobic system. Apart from the hydrogen ion concentration, several other compounds affect the rate of anaerobic digestion, even at very low concentration, such as heavy metals and chloro-organic compounds at inhibitory concentrations is unlikely in sewage.

2.2.8 Flow rate of the wastewater
Municipal wastewater is characterized by strong fluctuations in organic matter, suspended solids and flow rate. Concentrations of BOD, COD and TSS can vary with a factor of 2-10 in half an hour to a few hours (Mergeart et al., 1992). Flow rate fluctuations of domestic wastewater depend mainly on the size of population (the larger the population, the smaller the variation) and the sewer type (combined sewers have much higher fluctuations, due to receiving rain and run off water). Daily flow rate variations: the variation in flow tends to follow a diurnal pattern. The wastewater discharge curve closely follows the water consumption curve, but with a lag of several hours (Metcalf and Eddy, 2003).

2.3 Application of Anaerobic Treatment Technology for Municipal Wastewater

2.3.1 Perspectives of Anaerobic-Aerobic Systems:

In tropical countries sewage treatment, the aerobic processes (CAS (Conventional Activated Sludge) and MBR (Membrane Bioreactor) in the near future) have proven to be effective in producing high quality effluent to meet the discharge and water reclamation standards. However, aerobic systems are by nature, net energy consuming process, mainly due to the aeration requirements to sustain the aerobic microbial populations. Anaerobic process, on the other hand, does not require aeration and produces methane gas as a by-product during biodegradation of the complex organics, which can be utilized as fuel for energy production. Coupled by other advantages such as low sludge production, natural in process, simplicity in operation makes anaerobic technology environmentally friendly, cost-effective and economical.
2.3.2 Necessity of Aerobic Post-treatment Systems:

However, anaerobic processes are not very efficient when it comes to nutrients removal (such as nitrogen and phosphorus). Thus aerobic processes are still required as a polishing step on the anaerobic effluent to achieve the required standards for discharge or for further water reuse (Kobayashi et al., 1983). Whilst anaerobic processes have several advantages, it is important to realise that their treatment capacity is not sufficient. Treatment of simple organic material is reasonable, but for any additional treatment requirement, post-treatment processes are required. Therefore, in low- and middle-income countries where pollution should be of most concern, the use of anaerobic treatment in isolation is not sufficient.

Under “real life” conditions in developing countries, typical full scale process combinations (as presented in Figure 2.5) are however rarely entirely realised. Instead, often only the main treatment steps (aerobic wastewater treatment without a sludge digestion or anaerobic UASB treatment of sludge and wastewater without a post-treatment of the wastewater) are put in place in order to reduce the most severe environmental effects. Accordingly, post-treatment steps shown in Figure 2.5, below the dotted line are often not realised in developing countries as yet.
2.3.3 Assessment of Technological Requirements for Combined Systems:

Current technological limitations are the outcome of a failure to adjust to local conditions, experience and know-how, as well as the technology's short span of experience and development. This can be rectified by

- preliminary conclusion from the study – various anaerobic + post treatment coupled systems and their implications on cost related aspects can be done
- make study on the amount of sludge (%) that could be reduced (compared to existing system)
- assessment of energy reduction is needed
- the amount of biogas that would be produced and collected can be assessed
- the establishment and documentation of suitable examples of working plants
- the further development of the technology in terms of standardisation and cost-reduction measures
practical research and development in the areas of preliminary and post-treatment, pathogen removal emission and odour control, gas utilisation, sludge storage, small and medium-sized systems

- rehabilitation and improvement of existing plants

2.4 Progress of Anaerobic Treatment Technology for Municipal Wastewater

The first application of anaerobic digestion for sewage treatment is presumably the airtight chamber developed by the end of last century in France by M. Mouras (Van Hanndel and Lettinga, 1994). Around the change of the century, several new anaerobic treatment systems were developed. In 1935 world’s largest sewage treatment plant with imhoff tanks was constructed in Chicago (Van Hanndel and Lettinga, 1994).

In the following decades, anaerobic treatment of sewage became less popular than aerobic sewage treatment systems such as the trickling filter and activated sludge processes. This decreased application of anaerobic treatment was mainly due to higher removal efficiency of organic matter achieved in the aerobic systems. Well operated aerobic systems would remove 90 – 95 percent of the biodegradable organic matter from raw sewage. In the early anaerobic systems the removal was based on the settling of suspended organic matter. As only a fraction of the influent organic matter is settleable (one third to one half), the maximum removal efficiency in these systems did not exceed 30-50 percent of the biodegradable matter, depending on the nature of the sewage and the settling efficiency (Van Hanndel and Lettinga, 1994).

The low removal efficiency of the primary treatment systems must be attributed to a fundamental design failure. As there is little, if any, contact between the anaerobic micro-
organisms in the system and the non-settleable part of the organic matter in the influent, the main part of the dissolved or hydrolysed organic matter cannot be metabolized and leaves the treatment system. A very important aspect is the contact between the microorganisms and the wastewaters. The importance of a sufficient contact between influent organic matter and the bacterial population was not recognized at that time. The resulting relatively poor performance of anaerobic systems led to the belief that they were inherently inferior to aerobic systems, an opinion which often still persists today. However, in the mean time, it has been demonstrated that a properly designed modern anaerobic treatment system can attain a high removal efficiency of biodegradable organic matter, even at very short retention times.

A breakthrough in the design of anaerobic treatment systems came about with the development of ‘modern’ or high rate systems. All modern high rate anaerobic treatment systems are based on various kinds of sludge immobilization principle in order to retain as much sludge as possible. The different types of anaerobic treatment systems have been applied to a great variety of industrial wastes, but so far the anaerobic treatment concept is rarely used for sewage so experimental information is scarce. In fact, experimental results of anaerobic sewage treatment in modern systems are restricted to the use of the anaerobic filter (AF), fluidized and expanded bed (FB/EB) and upflow anaerobic sludge blanket (UASB). Comparison of process behaviour of all these three different modern high rate anaerobic processes is listed in Table 2.7.
Table 2.7 Comparison of different anaerobic process behaviour

<table>
<thead>
<tr>
<th>Characteristic behaviour</th>
<th>UASB</th>
<th>AF</th>
<th>FB/EB</th>
</tr>
</thead>
<tbody>
<tr>
<td>Reactor start-up</td>
<td>-</td>
<td>-</td>
<td>-</td>
</tr>
<tr>
<td>Biomass accumulation</td>
<td>*</td>
<td>+</td>
<td>*</td>
</tr>
<tr>
<td>Liquid-phase mixing</td>
<td>-</td>
<td>+</td>
<td>*</td>
</tr>
<tr>
<td>Robustness against hydraulic shocks</td>
<td>-</td>
<td>*</td>
<td>*</td>
</tr>
<tr>
<td>Robustness against organic shocks</td>
<td>+</td>
<td>+</td>
<td>+</td>
</tr>
<tr>
<td>Insensitivity to suspended solids</td>
<td>-</td>
<td>+</td>
<td>*</td>
</tr>
<tr>
<td>Insensitivity to clogging</td>
<td>*</td>
<td>-</td>
<td>*</td>
</tr>
<tr>
<td>Risk for biomass flotation</td>
<td>-</td>
<td>+</td>
<td>+</td>
</tr>
<tr>
<td>Demand for reactor control</td>
<td>+</td>
<td>+</td>
<td>-</td>
</tr>
</tbody>
</table>

- Unfavourable     + Favourable     * Very favourable

The upflow anaerobic filter system can suffer from clogging (channeling) problems. In UASB reactors, channeling problems occur only at low loading rates and when a poor feed-inlet distribution system has been installed in the reactor. In fluidized-bed reactors, a good contact between micro-organisms and wastewater is guaranteed and provided with a sophisticated feed-inlet distribution system. On the other hand, fluidized bed reactors require a high recycle factor, which may result in a distinct drop in substrate utilization rate by the active biomass because of the relatively low substrate levels prevailing in the reactor. In attached film processes, the maximum sludge retention depends mainly on the surface area for sludge attachment, the film thickness, the space occupied by the carrier material and the extent to which dispersed sludge aggregates are retained. In upflow
anaerobic filters the voidage of the packing material is a factor of prime importance with regard to sludge retention (Van Haandel and Lettinga, 1984).

### 2.5 Upflow Anaerobic Filter

Biofilm, or fixed film, reactors depend on the natural tendency of mixed microbial populations to adsorb to surfaces and to accumulate in biofilms. Adsorbed microorganisms grow, reproduce, and produce extracellular polymeric substances that frequently extend from the cell, forming the gelatinous matrix called a biofilm. Jimeno et al (1990) defined that bacterial attachment is mediated by polymeric material, primarily polysaccharide, which extended from the cell to form a tangled mass of fibers, termed a glycocalyx. The entire deposit is called a biofilm. The accumulation and persistence of a biofilm is the net result of several physical and biological processes that occur simultaneously, although their relative rates will change through the various stages. The mixing in these reactors is typical of plug flow (James and William, 1990).

In the upflow anaerobic packed-bed reactor the packing is fixed and the wastewater flows up through the interstitial spaces between the packing and biogrowth. While the first upflow anaerobic packed-bed processes contained rock, a variety of designs employing synthetic plastic packing are used currently. A large portion of the biomass responsible for treatment in the upflow attached growth anaerobic processes is loosely held in the packing void spaces and not just attached to the packing material (Metcalf and Eddy, 2003).
Anaerobic filter is filled out with a support material arranged in sheet, ring or sphere configuration which provides the best conditions for microbial attachment in biofilm form. The reactor may be operated in upflow or downflow mode (Bodik et al., 2000). In an upflow filter, the packing bed is fully submerged. The downflow can work either submerged or non-submerged. Process diagram of an upflow anaerobic filter is shown in Figure 2.6. The upflow anaerobic filter is basically a contact unit, in which wastewater passes through a mass of biological solids contained inside the reactor by a support medium. The biomass is contained in the reactor, by

1) biomass attached to the support media’s surface as a thin biofilm;
2) biomass entrapped within the media matrix; and
3) biomass held as a granulated or flocculated sludge mass beneath the media.

Figure 2.6 Schematic diagram of an upflow anaerobic filter.
Ramakrishnan and Gupta (2006) indicated that start-up of anaerobic reactors is more time consuming and is subjected to disturbances more than that of aerobic reactors. The start-up of the anaerobic process is still considered a major area of research. Many researchers have reported long start-up periods of 2 - 3 months to 1 year (or even more) for the anaerobic reactors. Accordingly, Punal et al. (2000) mentioned that long duration of start-up period is a major drawback of the anaerobic wastewater treatment systems.

Considerable efforts have been made to study the granulation process but the mechanism involved in the formation of granulation sludge is still unknown. A better understanding of the factors affecting biomass aggregation and adhesion, the two main mechanisms of biomass retention, could make the start-up more efficient and rapid. A feeding strategy, consisting of maintaining a low nitrogen concentration in the influent during the first two weeks followed by nitrogen balanced feed, is proposed in order to quicken the start-up of anaerobic filters. In addition, Subbiah (1997) reported that the start-up period required was about 54 days before the upflow anaerobic filter achieved steady state.

Low upflow velocities are generally used to prevent washing out the biomass as mentioned by Metcalf and Eddy (2003). Jimeno et al. (1990) reported that the C:N:P ratio of 100:2:1 is optimal for the start-up of anaerobic fixed-film reactors.

As the wastewater passes over the biomass, the soluble organic compounds contained in the influent wastewater, which is in contact with the biomass, are being diffused through the biofilm or the granular sludge. They are then converted into intermediate and final
products, specifically methane and carbon dioxide. The effluent from the anaerobic filter is usually well clarified and has relatively low concentration of organic matter.

Anaerobic filter can have several shapes, configurations and dimension, provided that the flow is well distributed over the bed. In full scale systems, anaerobic filters are usually present either in cylindrical or rectangular shape. The diameters of the tanks vary from 6 to 26 m and their heights from 3 to approximately 13 m. The volumes of the reactors vary from 100 to 10,000 m$^3$. Packing material may be in the entire depth or, for hybrid designs, only in the upper 50 to 70 percent (Van Hanndal & Lettinga, 1994).

### 2.5.1 Origin and Development of Anaerobic Filter

The first works on anaerobic filter dated back to the late 1960s and they have had a growing application since that time for treatment of both domestic wastewater and a diversity of industrial effluents. Table 8 shows the list of various anaerobic filters studied for different types of wastewater.

Two important developments in the application of anaerobic processes to lower strength wastewaters are the development of the anaerobic contact process by Schroepfer et al. (1955); Schroepfer and Ziemke (1959) and the development of the anaerobic filter by Coulter et al. (1957) and Young and McCarty (1968). The key concept of both processes relates to the ability to control mean cell retention time (MCRT) independently of hydraulic retention time. This feature permits anaerobic treatment at lower temperatures than previously thought possible or economical. Ng and Chin (1987) stated that without increasing MCRT independently of hydraulic retention time, very large reactor volumes
are required, making anaerobic treatment techniques too costly. Since heating is not required at tropical climate, low strength wastes, which produce only small quantities of gas per unit volume of waste treated, can be effectively treated by the anaerobic filter or anaerobic contact process. In addition, Kobayashi et al. (1983) stated that the filter performance at 25 and 35°C was not significantly different.

The modern anaerobic filter was reported as early as in 1968 by Young and McCarty (Van Hanndal & Lettinga, 1994). They reported a completely submerged, 12-L lab-scale reactor which was filled with 1.0 to 1.5 inch quartzite stone. The findings of Young and McCarty (1968) are as follows:

1) the anaerobic filter is ideal for the treatment of soluble wastewaters;
2) accumulation of biological solids in the anaerobic filter leads to long solids retention times (SRTs) and low effluent total suspended solids (TSS) and
3) low strength wastes were successfully treated at the temperature of 25°C because of long SRTs.

In addition to the initial studies done by Young and McCarty (1968), anaerobic filter was used to treat different types of wastewater by numerous researchers. Anaerobic filter is being used for treating high strength industrial wastewater for a long time. Ng and Chin (1987) had used a lab-scale anaerobic filter to treat piggery wastewater successfully. And Herbert et al. (1994) studied a lab-scale hybrid system of UASB and anaerobic filter to treat synthetic wastewater comprising milk and sucrose with balanced nutrients and trace metals. It was reported that a hybrid system of UASB and anaerobic filter could achieve
95% of COD removal, which was higher than that achieved in an expanded bed and fluidized bed reactors. Bodik et al. (2002) studied the feasibility of anaerobic sequencing batch reactor (AnSBR) and anaerobic filter reactors to treat synthetic and domestic wastewater. They concluded that AnSBR and upflow anaerobic filter seem to be potential options for pre-treatment of wastewater produced by small communities. Lab scale and pilot scale plants which are reviewed in literature study are listed in Table 2.8.

Kobayashi and his coworkers (1983) used a lab-scale anaerobic filter packed with synthetic high surface area trickling filter media to treat low strength domestic wastewater at temperatures of 20, 25 and 35°C at a HRT of 24 hours. From their study it was concluded that the anaerobic filter is a promising process for treatment of low strength wastewaters, and that post-treatment for sulfides and ammonia removal may be necessary. In 1998, Chernicharo and Machado evaluated the applicability of a pilot-scale anaerobic filter for polishing domestic sewage after its pre-treatment by an UASB. The performance of upflow and downflow anaerobic filters was compared and they concluded that the overall performance of the upflow anaerobic filter was better than the down flow anaerobic filter.
Table 2.8. List of reviewed plants in various studies

<table>
<thead>
<tr>
<th>Type of Wastewater</th>
<th>Scale and Temperature (°C)</th>
<th>OLR (l/1000/day)</th>
<th>COD removal Efficiency (%)</th>
<th>HRT (hrs)</th>
<th>Media Type</th>
<th>Methane Composition (%)</th>
<th>Reference</th>
</tr>
</thead>
<tbody>
<tr>
<td>Domestic</td>
<td>Lab scale UAF 20, 25, 30</td>
<td>0.02 lbCOD ft⁻³ day⁻¹</td>
<td>73</td>
<td>24</td>
<td>Trickling filter media</td>
<td>92 to 98% (ignoring nitrogen content)</td>
<td>Kobayashi et al., (1983)</td>
</tr>
<tr>
<td>Piggery wastewater</td>
<td>Lab scale UAF 30</td>
<td>5 g/L.d</td>
<td>97 – 83%</td>
<td>6.3 to 2.1 days</td>
<td>PVC tubes</td>
<td>75 to 84%</td>
<td>Ng et al., (1987)</td>
</tr>
<tr>
<td>Currant-finishing wastewater</td>
<td>Pilot Scale DAF (80 &amp; 5000 liters) 35±1</td>
<td>1 &amp; 8 kg COD/m3 d</td>
<td>&lt;80 and 80%</td>
<td></td>
<td>Pall rings &amp; modular corrugated cross-flow pieces</td>
<td></td>
<td>Athanasopoulos et al., (1990)</td>
</tr>
<tr>
<td>Synthetic wastewater</td>
<td>Lab Scale Hybrid UASB+UAF 37</td>
<td>20 g/L.d</td>
<td>95%</td>
<td>3</td>
<td>Plastic tubes</td>
<td></td>
<td>Herbert et al., (1994)</td>
</tr>
<tr>
<td>Saline wastewater</td>
<td>Lab scale 37</td>
<td>5 kg COD/m3.d with 7.5g Cl/l</td>
<td>80%</td>
<td></td>
<td>PVC rings</td>
<td></td>
<td>Guerrero et al., (1997)</td>
</tr>
<tr>
<td>Wastewater from coffee processing plant</td>
<td>Pilot scale</td>
<td>Domestic Lab scale</td>
<td>Multistage UAF</td>
<td>Lab scale &amp; Municipal wastewater</td>
<td>Lab scale 24, 15, 9</td>
<td>Domestic Lab scale 24+1</td>
<td>Synthetic Substrate 24</td>
</tr>
<tr>
<td>---------------------------------------</td>
<td>------------</td>
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<td>----------------</td>
<td>---------------------------------</td>
<td>--------------------</td>
<td>------------------------</td>
<td>------------------------</td>
</tr>
<tr>
<td>4 days to 8h</td>
<td>Waste tyre tube</td>
<td>Plastic fillings</td>
<td>Reticulated Polyurethane Foam (RPF)</td>
<td>Plastic material</td>
<td>Reticulated Polyurethane Foam (RPF)</td>
<td>63</td>
<td>81</td>
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<tr>
<td>97 to 86%</td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td>64</td>
<td>87</td>
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<td>90</td>
<td>66</td>
<td>63</td>
<td>77</td>
<td>81</td>
<td>70.7±2.9</td>
</tr>
</tbody>
</table>
In addition, Athanasopoulos et al. (1990) studied two down flow anaerobic filters with different plastic media of the same specific area, treating currant-finishing wastewater and concluded that down flow anaerobic filters had a lower performance compared with other high-rate anaerobic reactors, UAF and UASB.

Bodik et al. (2000) studied a lab-scale upflow anaerobic filter and pilot-scale anaerobic baffled filter to treat municipal wastewater and their research findings confirmed that anaerobic wastewater treatment process was suitable for municipal or domestic wastewater and the pilot-scale reactor worked during the whole experiments without any technological problems; no significant changes of pH, VFA were observed in the anaerobic reactor.

In addition, Fatma and Michael (2003) developed a dynamic mathematical model to understand the applicability of anaerobic treatment for low strength wastewater. The model had served as a predictive tool for treatment efficiency and gas production.

**Advantages of Upflow Anaerobic Filter**

Ng et al. (1987) stated that anaerobic processes are usually limited by the low growth rate of the methanogens. Due to this limitation, conventional suspended-growth anaerobic treatment systems require lengthy retention times and thus large reactor volume. The advantages of anaerobic filter (an attached-growth system) over a suspended-growth anaerobic high-rate reactor are as follows:
1) Biofilm reactors are especially useful when slow growing organisms have to be kept in wastewater treatment (Bodik et al., 2003).

2) It has relatively good load fluctuation resistance (Kobayashi et al., 1983; Nebot el al., 1995; Bodik et al., 2000; Francisco Omil et al., 2003).

3) In anaerobic filter, bacteria adhere to support media so that, even at relatively high hydraulic loads (which would wash bacterial biomass out of conventional suspended growth digesters), the filter retains the bacteria (Young and McCarty, 1968).

4) The amount of produced sludge is smaller and settleability of sludge is good (Bodik et al., 2000).

5) Due to the efficient biomass retention, long sludge ages and more compact reactors can easily be achieved (Kobayashi et al., 1983; Bodik et al., 2003).

6) Sludge is not returned, unlike the anaerobic activated sludge process (Bodik et al., 2000). Therefore cost of energy for sludge returning is not necessary.

7) Suitable for treatment of low soluble organic wastewater (Bodik et al., 2000).
**Disadvantages of Upflow Anaerobic Filter**

However, anaerobic filter has some drawbacks too. The disadvantages of anaerobic filter are as follows (Bodik *et al.*, 2000):

1) channeling can occur, i.e. formation of preferential paths of liquid flow through the reactor.

2) dead-zone formation caused by sludge compaction or clogging of matrix interstitial spaces by solids.

3) clogging of poorly designed distribution systems.

**Packing Media**

The purpose of packing medium is to retain solids inside the reactor, either by the biofilm formed on the surface of the packing medium or by the retention of solids in the interstices of the medium or below it. The main purposes of the packing media are as follows:

1) acting as a device to separate solids from liquid;

2) helping to promote a uniform flow in the reactor;

3) improving the contact between the components of the influent wastewater and the biological solids contained in the reactor;

4) allowing the accumulation of high amount of biomass, with a consequently increased solids retention time; and

5) acting as a physical barrier to prevent solids from being washed out from the treatment system.
Several types of materials have been used as packing media in biological reactors, including quartz, ceramic blocks, oysters and mussel shells, limestone, plastic rings, hollow cylinders, PVC modular blocks, granite, polyethylene balls and bamboo. The packing media have been designed to occupy from the total depth of the reactor to approximately 50 to 70% of the height of the reactor. There are different types of plastic packing media available in the market, ranging from corrugated rings to corrugated plate blocks. The specific surface areas of these plastic materials usually range from 100 to 200 m$^2$/m$^3$. Although some types of packing media are more efficient than others in the retention of biomass, the final choice will depend on the specific local conditions, economic considerations and operational factors. The requirements for good packing media of anaerobic filter are listed in Table 2.9.

Elmitwalli et al. (2000) indicated that specific surface area, porosity, surface roughness, pore size and orientation of the packing material were important factors influencing the anaerobic filter reactor performance. High surface area and porosity, large pore size and rough surface area for packing material improved performance of an AF reactor. Subsequently, Mohammad (2000) stated if an excessively small medium is employed AFs may suffer from blockages and to minimize blockages, filter media tend to have relatively large diameters (>20 mm). The surface roughness of packing filter media and degree of porosity, in addition to pore size, affect the rate of colonization by bacteria (Stronach et al., 1986).
<table>
<thead>
<tr>
<th>Requirement</th>
<th>Objective</th>
</tr>
</thead>
<tbody>
<tr>
<td>Structural resistance</td>
<td>Support their own weight and the weight of the biological solids attached to the surface</td>
</tr>
<tr>
<td>Biological and chemical inertness</td>
<td>Allow no reaction between the bed and the microorganisms</td>
</tr>
<tr>
<td>Sufficient light</td>
<td>Avoid the need for expensive, heavy structures, and allow the construction of relatively higher filters, which implies a reduced area necessary for the installation of the system</td>
</tr>
<tr>
<td>Large specific area</td>
<td>Allow the attachment of a larger quantity of biological solids</td>
</tr>
<tr>
<td>High porosity</td>
<td>Allow a larger free area available for the accumulation of bacteria and reduce the possibility of clogging</td>
</tr>
<tr>
<td>Enable the accelerated colonization</td>
<td>Reduce the start-up time of the reactor</td>
</tr>
<tr>
<td>of microorganisms</td>
<td></td>
</tr>
<tr>
<td>Present a rough surface and a non-flat format</td>
<td>Ensure good attachment and high porosity</td>
</tr>
<tr>
<td>Low cost</td>
<td>Make the process feasible, not only technically, but also economically</td>
</tr>
</tbody>
</table>
CHAPTER 3 MATERIALS AND METHODS

3.1 Lab Scale Upflow Anaerobic Filter

3.1.1 Experimental Setup of Anaerobic Filters

Two cylindrical upflow anaerobic filters - UAF1 and UAF2, were constructed from acrylic plates and columns. Figures 3.1 and 3.2 show the picture and schematic diagram of the experimental setup of the UAF system, respectively. The effective volume of UAF1 and UAF2 were 20.5 and 17.8 L, respectively. To solve the problem experienced for measuring the attached growth biomass in UAF1, the UAF2 reactor was constructed with an acrylic column of diameter 0.085 m and height 1.5 m and located at the centre of the UAF2. Therefore, the effective volume of UAF2 was reduced slightly. The diameters of both reactors were 0.14 m, and height of UAF1 and UAF2 were 1.67 and 1.72 m, respectively.

Both reactors were filled with PVC medium (Sera Siprox D52518, Aquaristic, Germany), which has a length of 25 mm and a diameter of 12 mm. This type of medium has the capacity to provide about 270 m$^2$ effective surface area per 1 L of medium. The heights of the filtration medium of UAF1 and UAF2 were 1.30 and 1.40 m, respectively. The whole experimental set-up consisted of raw sewage tank, sewage transfer tank, anaerobic filter, effluent tank and biogas collector. Although provision was made for desludging the filter, it would be prudent to incorporate this facility in any full-scale plant.
Figure 3.1 Picture of Upflow Anaerobic Filters
Figure 3.2 Schematic diagram of UAF experimental set up
3.1.2 Seeding

Both UAF1 and UAF2 reactors were seeded with 16.3 L of screened anaerobic digester sludge from the top of the reactor. The size of the sieve was 2 mm. Anaerobic sludge was obtained from anaerobic digester of Ulu Pandan Water Reclamation Plant (UPWRP) treating domestic wastewater. Mixed liquor suspended solids (MLSS) and mixed liquor volatile suspended solids (MLVSS) concentrations of anaerobic sludge were 10,833 mg/L and 9700 mg/L, respectively. The reactors were fed with domestic wastewater. Nitrogen gas was passed several times into the reactors to replace the oxygen present inside the reactor. Recirculation of sludge within the reactors was done for first 2 days to allow even distribution of biomass. Long retention times of 36 and 24 hrs were maintained for first two weeks to allow the biomass to grow and attach on medium while avoiding biomass washout.

3.1.3 Operating Conditions

Domestic wastewaters were also collected from Ulu Pandan Water Reclamation Plant (UPWRP). The wastewater was stored in a cold room at 4°C to reduce degradation during storage. The wastewater was fed into the sewage feed tank daily after screening with a sieve of 2 mm aperture in order to avoid clogging of filter by bigger particles. The wastewater was then transferred to the feed transfer tank. A heater was provided in the sewage transfer tank to maintain a temperature of 30°C in the influent, which is ambient temperature in tropical countries. From the influent transfer tank, domestic wastewater was introduced into the filters at the bottom of the reactor with the help of peristaltic pumps. Magnetic stirrers were used at the bottom of both reactors to avoid settling of
suspended solids present in the domestic wastewater. Two pH probes were provided to monitor the pH of effluent from the reactors in order to maintain the pH between 6.8 and 7.2. Since higher concentrations of Volatile Fatty Acids (VFA) accumulation was expected in the reactors, stand by alkaline dosage arrangement was set up to increase the pH. A biogas collector was provided for each UAF to collect the biogas produced from each reactor. They are floating covered gas collectors with counter-weights. The water inside the biogas cylinder containers were acidified to pH of around 4. There was no effluent recirculation in both UAFs and no intended sludge wasting throughout the operation. The operating conditions maintained for the anaerobic filters were as follows:

\[
\begin{align*}
\text{pH} & \quad -6.8 \text{ to } 7.2 \\
\text{Temperature} & \quad -30^\circ C \\
\text{HRTs} & \quad -24, 16, 12, 8, 6, \text{ and } 4\text{hrs}
\end{align*}
\]

Influent, effluent, biomass and biogas samples were collected from the reactors and analyzed.

### 3.2 Analytical Methods

Performance of anaerobic filters were monitored by analyzing suspended solids (SS), volatile suspended solids (VSS), total chemical oxygen demand (tCOD), soluble chemical oxygen demand (sCOD), five-day total biochemical oxygen demand (tBOD\(_5\)), five-day soluble biochemical oxygen demand (sBOD\(_5\)), alkalinity, volatile fatty acids (VFAs), anions, cations of influent and effluent, mixed liquor suspended solids (MLSS), mixed liquor volatile suspended solids (MLVSS), gas production and gas composition.
These parameters were tested in accordance with the Standard Methods listed for water & wastewater (APHA, 1998).

3.2.1 Total Suspended Solids (TSS) and Volatile Suspended Solids (VSS)

Total suspended solids and volatile suspended solids concentrations were determined according to the method specified in the Standard Methods (APHA, 1998). For TSS measurement, the sample was dried in an oven (MEMMERT ULM 6, Schmidt Scientific) at 105°C for 1 hour. For VSS measurement, the sample was further burned in a furnace (Thermolyne 48000, Omega Medical Scientific) at 550°C for 20 minutes.

3.2.2 Chemical Oxygen Demand (COD)

The closed reflux method (Block heater: HACH COD Heater, Model 16500-10) in accordance with the Standard Methods (APHA, 1998) was used to measure COD.

3.2.3 Biochemical Oxygen Demand (BOD)

The BOD measurements were done in accordance with the Standard Methods (APHA, 1998). The dissolved oxygen (DO) concentration in the samples was monitored with a DO meter (YSI Model-58, YSI Integrated Systems & Services, United States of America).
3.2.4 Biogas Composition

The biogas composition was measured using the gas chromatography (Shimadzu GC-17A, Shimadzu Scientific Instruments, United States of America) unit with packed column (80/100 PORAPAK, 2m X 1/8 in., SUPELCO, United States of America) using Argon as the carrier gas. Calibration was performed using 100 µl of standard gas comprising 25% H₂, 25% N₂, 10% CH₄ and 40% CO₂. One hundred microliter of sample was injected in each run and the experiment was repeated 3 times.

3.2.5 Total Nitrogen (TN)

TN concentrations of the samples were measured using Shimadzu TOC analyzer (Shimadzu TOC-VCSH, Shimadzu Scientific Instruments, United States of America) and a TN measuring unit (Shimadzu TNM-1, Shimadzu Scientific Instruments, United States of America).

3.2.6 Total Phosphorus (TP)

The sample was first pre-treated using test kit from HACH (total phosphate, 0-3.5 mg/L, reagent set 27426-45, HACH Company, United States of America). The total phosphorus concentration was then analyzed with a spectrophotometer (HACH DR/4000U, HACH Company, United States of America).
3.2.7 Volatile Fatty Acid (VFA)

The Flame Ionization Detector (FID) (Shimadzu GC-2010, Shimadzu Scientific Instruments, United States of America) with capillary column (25 m x 0.32 mm, HP-FFAP, Agilent Technologies Inc., United States of America) was used to detect VFA concentration of the influent sewage and anaerobic treated effluent samples. Acetic, propionic, butyric and n-valeric acid were prepared as standards for calibration up to concentrations of 200 mg/L. Nine parts of sample was mixed with 1 part of formic acid before injecting into the unit for analysis. An injection volume of 0.2 ml at column temperature of 150°C was used and run time for each sample was 10 minutes.

VFAs were analysed using a gas chromatograph (GC, Chrompack CP 9000, Varian, Inc. Scientific Instruments, United States of America) equipped with a flame ionization detector. The VFA analytic procedure was in accordance with that described by Buyukkamaci et al. 2004. The volatile acids in the sample were separated and quantified as free acids by using a capillary column (Nukol, 30 m X 0.25 mm i.d., 0.25um film; Supelco Cat. No: 4-6875, SUPELCO, United States of America). Nitrogen gas was used as carrier gas with a rate of 20 cm s\(^{-1}\). Temperature conditions were: column temperature limit, 2500C; oven initial, 1400C; and oven final, 1850C. Split ratio was 100:1 at 2200C. Before starting analysis of the samples taken from the reactor, calibration curves for acetic, propionate, butyric and valeric acids were prepared with the samples containing known concentrations of the volatile fatty acid of the interest. Any suspended solids were removed before injecting into the column to prevent any clogging in the GC. Therefore,
samples were centrifuged and filtered through a membrane filter (0.45µm), in order to obtain suspended solids-free content samples.

3.2.8 Ammonia-Nitrogen (NH$_4^+$-N)

NH$_4^+$-N was measured by using the 4500-H Automated Phenate Method with the Mark III multi-channel color meter continuous flow analysis setup (Auto Analyser Accessories, Chemlab Instrument, United Kingdom) in accordance with Standard Methods (APHA, 1998).

3.2.9 Total Organic Carbon (TOC)

Total Organic Carbon Analyzer (TOC-VCSH, Shimadzu Scientific Instruments, United States of America) with auto sampler and injector (ASI-V, Shimadzu Scientific Instruments, United States of America) was used to determine the organic carbon concentration of the samples. All samples were diluted to less than 25 mg/L before analysis. The method used was 680°C catalytically-aided combustion oxidation.

3.2.10 Anion Concentrations

Anion concentrations in the soluble portion of the wastewater sample were measured by ion chromatography (Dionex Corporation, United States of America) using Dionex® AS9-HC analytical column (Dionex Corporation, United States of America). Sample tubes were pre-washed with distilled water and sonicated (NEY ULTRASONIK, Equivalent Scientific Products, United States of America) for 30 minutes. Twenty milli-liter of sample was injected into the column and eluted with 0.009 M of sodium carbonate.
3.2.11 Hydrogen Sulphide (H$_2$S)

The H$_2$S content of biogas was measured by gas chromatography (Shimadzu GC-2010, Shimadzu Scientific Instruments, United States of America) using the Flame Photometric Detector (FPD) and capillary column (25m x 0.32mm, GS-CASPRO, J&W Scientific Inc., United States of America). Calibration was done with standard gas comprising of 90% N$_2$ and 10% H$_2$S. For each biogas sample, 1 µl of biogas was injected and the experiment was repeated 3 times.

3.2.12 Alkalinity

The alkalinity was measured by titration according to the method specified in the Standard Methods (APHA, 1998) using 0.1 N of hydrochloric acid and an automated-titrator (Metrohm Titrando 808, Metrohm, United States of America).

3.2.13 Molecular Weight (MW) Distribution

MW distribution were determined using a 50 ml stirred ultrafiltration cell (Amicon® model 8050, Millipore Corporation, USA) using 44.5 mm Millipore disc ultrafiltration membranes. Three membranes with nominal MWs of 100,000 (100K), 10,000 (10K) and 1,000 (1K) daltons were used in succession with the highest MW first and lowest MW last. Pure nitrogen was used to pressurize the cell. The pressure in the ultrafiltration cell
was kept constantly at 30 psi during the sample filtration. Samples taken after each of the filters were analysed to determine the specific total organic carbon (TOC) (TOC-VCSH, Shimadzu Scientific Instruments, United States of America).

### 3.2.14 EPS Extraction

Extracellular polymers (EPS) were extracted by a combination of sonication and cation exchange resin treatment (CER) (Dignac *et al.*, 1998). Extracellular polymeric substance (EPS) was separated from the microorganism cell wall by using cation resin exchange. Cation exchange resin (CER) will remove cations from the sludge matrix leading to a break up of the flocs and a subsequent release of EPS. The CER was firstly washed in phosphate buffer and stirred for an hour. Thereafter, the phosphate buffer was changed. The EPS extraction procedure is:

1. 75 ml of the sludge sample was centrifuged for 10 minutes at 9,000 rpm (4°C);
2. The supernatant was decanted and resuspended to the original volume using phosphate buffer;
3. 70 g CER/g VSS was then added to the suspension in an open-mouth closed container;
4. The suspension was stirred at 600 rpm for 1.5 hours in the cold room (4°C);
5. The suspension was then centrifuged at 9,000 rpm for 10 minutes to separate the CER and biomass; and
(6) The supernatant was finally collected for subsequent analysis of EPS for proteins and carbohydrates.

Extracellular polymers (ECP) are known to play a key role in wastewater treatment: they are important for the removal of pollution from wastewater, and for sludge settling Dignac et al. (1998). Even where some relationships between ECP and process performance have been described, comparisons are difficult due to the diversity of sludge samples, the variety of extraction methods, and to a large extent due to the diversity and lack of confidence in the chemical analysis methods.

**Proteins**

The method described by Lowry et al. 1951 was followed except for some slight modifications in the preparation of reagents. The first step is a biuret reaction where peptide bonds in protein react with copper in alkaline solution. The next step is a reduction of the active phosphomolybdic and phosphotungstic acids in the reagent by the copper treated protein. The colour developed was measured spectrophotometrically at an absorbance of 650 nm using a UV-Vis spectrophotometer (HACH DR/4000, HACH Company, United States of America) to determine the concentration of proteins in the biomass.
Carbohydrates

The procedure described by Dubois et al. (1956) was followed using the phenol reagent as a 5% solution in water. The sample was heated with strong sulphuric acid together with the reagent to develop an orange colour. The sample was then measured spectrophotometrically at an absorbance of 490 nm using a UV-Vis spectrophotometer (HACH DR/4000, HACH Company, United States of America).
CHAPTER 4 RESULTS AND DISCUSSION – UAF

UAF1 and UAF2 were commissioned on 12 May 2005 and 14 Sep 2005, respectively. The experimental study was to investigate their long-term performance at different hydraulic retention times (HRTs). UAF1 was operated at a HRT of 16, 8 and 4 hours, while UAF2 was operated at a HRT 12, 6 hours and fluctuating HRTs.

Mergeart et al. (1992) demonstrated that municipal wastewater was characterized by strong fluctuations in organic matter, suspended solids and flow rate. Concentrations of BOD, COD and TSS varied with a factor of 2-10 in half an hour to a few hours. Flow rate fluctuations in domestic wastewater depend mainly on the size of population (the larger the population, the smaller the variation) and the sewer type (combined sewers have much higher fluctuations, due to receiving rain and run off water). Daily flow rate variations (the variation in flow) tends to follow a diurnal pattern. Based on observed results for domestic wastewater, the total loading of waste was noted to remain about the same on a daily basis throughout the year, even though the flow rate and the concentration vary. In this research fluctuating HRTs, simulating the normal fluctuating household wastewater discharged to the treatment plant, were conducted to study the effect of fluctuating loading on the performance of the UAF. The daily fluctuating operation mode was as follows:

10 am – 10 pm @ HRT 4 hours sampling around 9 pm everyday
10 pm – 10 am @ HRT 6 hours sampling around 9 am everyday
Parameters such as TSS, VSS, tCOD, sCOD, tBOD, sBOD, biogas production and biogas composition were monitored at each operating HRT to study the performance of the reactors.

### 4.1 Suspended Solids Removal

The influent and effluent suspended solids (SS) concentrations and removal efficiencies of UAF1 and UAF2 at different HRTs are shown in Figures 4.1 and 4.2. Table 4.1 summarizes the range and average influent and effluent suspended solids (SS) concentrations and removal efficiencies under different hydraulic retention times.

Anaerobic bacteria are slower growing microorganisms compared with the aerobic bacteria (Metcalf & Eddy, 2003). Moreover, methanogenic bacteria are slower growing bacteria compared with the acid forming bacteria. Therefore, HRT of more than 1 day was applied during the initial start-up of UAFs for about 18 days to facilitate methanogenic bacteria to form biofilm.

Solids removal efficiency can be affected by temperature, organic loading rate (OLR), hydraulic retention time (HRT) and upflow velocity (Mahmoud et al., 2003). During the startup period, HRT of 16 hrs was maintained for about 221 days in UAF1. Under this HRT, the influent SS concentrations ranged from 205 to 868 mg/L, with an average of 410 mg/L and the effluent SS concentrations ranged from 22 to 268 mg/L, with an average of 86 mg/L. There was much variability in feed quality due to reasons such as
high rainfall, which would dilute the feed. At 16 hours HRT, periodical biomass washout was observed every 30 to 40 days interval. This phenomenon in turn caused the removal efficiencies of SS and VSS to deteriorate to below 50%. Following this, the reactor performance would recover after 3 to 4 days. The reason for these biomass washouts could be due to the accumulation of SS inside the reactor, since no sludge wasting was carried out. The observed SS removal efficiencies ranged from 32 to 97 %, with an average value of 78%.

UAF2 was operated at 12 hours HRT for about 75 days. The observed results in UAF2 were relatively unstable. There was no apparent periodical biomass washout observed at 12 hours HRT. The reason might be due to the shorter HRT which led to a higher effluent solids concentration (more biomass was continuously wash out with the effluent). This in turn reduced the chances of periodical biomass washout. The effluent suspended concentration at 12 hours HRT was slightly higher than that at 16 hours HRT due to high hydraulic loading rate at a shorter HRT. Subsequently, 4 % reduction in the SS removal efficiency was noted at 12 hrs HRT compared with 16 hrs HRT. On Day 76, the connection between influent tube and reactor nozzle was accidentally detached. All biomass inside the reactor was washed out from the influent port. Therefore, the reactor was reseeded on Day 95 and was closely monitored.
Table 4.1. Influent and Effluent Suspended Solids Concentration and Removal Efficiencies

<table>
<thead>
<tr>
<th>Characteristics</th>
<th>HRT 16 hrs (UAF1)</th>
<th>HRT 12 hrs (UAF2)</th>
<th>HRT 8 hrs (UAF1)</th>
<th>HRT 6 hrs (UAF2)</th>
<th>HRT 4 hrs (UAF1)</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>Range</td>
<td>Range</td>
<td>Range</td>
<td>Range</td>
<td>Range</td>
</tr>
<tr>
<td>Influent SS (mg/L)</td>
<td>205 ~ 868</td>
<td>258 ~ 724</td>
<td>220 ~ 1030</td>
<td>220 ~ 1030</td>
<td>320 ~ 1057</td>
</tr>
<tr>
<td></td>
<td>Av</td>
<td>Ave</td>
<td>Ave</td>
<td>Ave</td>
<td>Ave</td>
</tr>
<tr>
<td>Effluent SS (mg/L)</td>
<td>22 ~ 268</td>
<td>67 ~ 263</td>
<td>33 ~ 252</td>
<td>40 ~ 232</td>
<td>82 ~ 359</td>
</tr>
<tr>
<td></td>
<td>86</td>
<td>103</td>
<td>128</td>
<td>133</td>
<td>169</td>
</tr>
<tr>
<td>SS Removal (%)</td>
<td>32 ~ 97</td>
<td>30 ~ 90</td>
<td>50 ~ 93</td>
<td>53 ~ 94</td>
<td>50 ~ 84</td>
</tr>
<tr>
<td></td>
<td>78</td>
<td>74</td>
<td>74</td>
<td>73</td>
<td>66</td>
</tr>
</tbody>
</table>
Following this, anaerobic filter (UAF2) was operated at 8 hours HRT for about 288 days. The effluent SS concentrations at 8 hours HRT were higher than that at 16 and 12 hours HRT owing to high hydraulic loading rate at a shorter HRT. The possible reason could be that the UAF2 was directly operated at 8 hrs HRT after the wash out incident stated earlier.

The UAF2 was later operated at 6 hours HRT for about 198 days. The effluent SS concentration at 6 hours HRT was higher than that at 16, 12 and 8 hours HRTs. Subsequent shortening the HRT to 4 hours demonstrated further deterioration in the SS removal efficiency by 7% compared with 6 hours HRT. Henz and Harremoes (1982) explained that the cause for biomass wash out of anaerobic filters are due to gas bubbles adhered to flocs/bed particles and caused them to rise in the reactor, eventually leading to washout of biomass and deterioration of the effluent quality.
Figure 4.1. Influent and effluent SS and VSS concentrations and removal efficiencies of UAF1 at HRT 16, 8 and 4 hours
Figure 4.2. Influent and effluent SS and VSS concentrations and removal efficiencies of UAF2 at HRT 12 and 6 hours
Finally, UAF2 was operated under the fluctuating HRTs of 6 hours and 4 hours for about 178 days. Each HRT was operated for 0.5 day. Influent and effluent SS concentrations and removal efficiencies of UAF2 at fluctuating HRT are shown in Figure 4.3. Table 4.2 shows the range and average influent and effluent SS concentration and removal efficiencies at fluctuating loading rate.

Table 4.2. Influent and Effluent Suspended Solids Concentrations and Removal Efficiencies of UAF2 at fluctuating loading rates

<table>
<thead>
<tr>
<th>Characteristics</th>
<th>HRT = 6 hrs</th>
<th>HRT = 4 hrs</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>Range</td>
<td>Average</td>
</tr>
<tr>
<td>Influent, SS (mg/L)</td>
<td>230 ~ 1057</td>
<td>437</td>
</tr>
<tr>
<td>Effluent, SS (mg/L)</td>
<td>60 ~ 404</td>
<td>156</td>
</tr>
<tr>
<td>SS % reduction</td>
<td>7 ~ 88</td>
<td>63</td>
</tr>
</tbody>
</table>

During this operation, the influent SS concentrations ranged from 230 to 1057 mg/L, with an average of 437 mg/L and the effluent SS concentrations for 6 hours HRT and 4 hours HRT were from 60 to 404 mg/L, with an average of 156 mg/L and from 50 to 530 mg/L, with an average of 164 mg/L, respectively. There was insignificant difference in the average SS removal efficiency at 4 hours and 6 hours HRTs. This observation agreed with Kobayashi et al. (1983) who stated that removal efficiencies showed very little sensitivity to daily fluctuations in influent domestic wastewater quality.
Figure 4.3 Influent and effluent SS concentrations and removal efficiencies of UAF2 at fluctuating HRT of 6 and 4 hours
From the above results, it can be concluded that shorter HRT would lead to higher SS concentrations in the effluent and corresponding lower SS removal efficiencies. And, based on SS removal, the optimum HRT for treatment of domestic wastewater by the upflow anaerobic filter was found to be 6 hrs.

4.2 Volatile Suspended Solids (VSS) Removal

The influent and effluent VSS concentrations and removal efficiencies of UAF1 and UAF2 at different HRTs are shown in the Figures 4.1 and 4.2. Table 4.3 summarizes the range and average influent and effluent VSS concentrations and their removal efficiencies under different HRTs.

While the influent VSS concentrations were ranged from 165 to 767 mg/L, with an average value of 332 mg/L, the effluent VSS concentrations were ranged from 17 to 217 mg/L, with an average value of 69 mg/L at 16 hours HRT. The removal efficiencies of VSS ranged from 26 to 97%, while the average removal efficiency was at 77%.

The effluent VSS concentration at 12 hours HRT was slightly higher than that at 16 hours HRT. Higher effluent VSS concentration led to 4% reduction in the VSS removal efficiency compared with that at 16 hours HRT. This was due to the high hydraulic loading rate in a shorter HRT as stated earlier.
<table>
<thead>
<tr>
<th>Characteristic</th>
<th>HRT 16 hrs (UAF1)</th>
<th>HRT 12 hrs (UAF2)</th>
<th>HRT 8 hrs (UAF1)</th>
<th>HRT 6 hrs (UAF2)</th>
<th>HRT 4 hrs (UAF1)</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>Range Ave</td>
<td>Range Ave</td>
<td>Range Ave</td>
<td>Range Ave</td>
<td>Range Ave</td>
</tr>
<tr>
<td>Influent VSS  (mg/L)</td>
<td>165 ~ 767 332</td>
<td>210 ~ 491 329</td>
<td>197 ~ 793 427</td>
<td>197 ~ 684 445</td>
<td>183 ~ 555 371</td>
</tr>
<tr>
<td>Effluent VSS (mg/L)</td>
<td>17 ~ 217 69</td>
<td>60 ~ 161 84</td>
<td>20 ~ 244 110</td>
<td>31 ~ 224 114</td>
<td>62 ~ 250 128</td>
</tr>
<tr>
<td>VSS Removal (%)</td>
<td>26 ~ 97 77</td>
<td>34 ~ 87 73</td>
<td>50 ~ 92 73</td>
<td>52 ~ 93 72</td>
<td>39 ~ 82 64</td>
</tr>
</tbody>
</table>
The effluent VSS concentration at 8 hours HRT was also higher than that at 16 and 12 hours HRT. Even though 8 hours HRT resulted in higher effluent VSS concentration, the VSS removal efficiency was similar to that at 12 and 16hrs HRTs. Besides, the average VSS removal efficiency at HRT 6 hours were insignificantly lower than that at 16, 12 and 8 hours HRTs. Likewise, the effluent VSS concentration at 4 hours HRT was higher than that at 16, 12, 8 and 6 hours HRT which leaded to 8% reduction in VSS removal efficiency at 4 hrs compared with 6 hrs HRT.

Influent and effluent VSS concentrations and removal efficiencies of UAF2 at fluctuating HRT were shown in Figure 4.4. Table 4.4 shows the range and average influent and effluent VSS concentration and removal efficiencies at fluctuating loading rate.

**Table 4.4. Influent and Effluent Volatile Suspended Solids Concentrations and Removal Efficiencies of UAF2 at Fluctuating Loading Rate**

<table>
<thead>
<tr>
<th>Characteristics</th>
<th>HRT = 6 hrs</th>
<th>HRT = 4 hrs</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>Range</td>
<td>Average</td>
</tr>
<tr>
<td>Influent, VSS (mg/L)</td>
<td>183 ~ 593</td>
<td>360</td>
</tr>
<tr>
<td>Effluent, VSS (mg/L)</td>
<td>59 ~ 304</td>
<td>129</td>
</tr>
<tr>
<td>VSS % reduction</td>
<td>14 ~ 88</td>
<td>63</td>
</tr>
</tbody>
</table>
The influent VSS concentrations were ranged from 183 to 593 mg/L, with an average of 360 mg/L and the effluent VSS concentrations for 6 hours HRT and 4 hours HRT ranged from 59 to 304 mg/L, with an average of 129 mg/L and from 39 to 445 mg/L, with an average of 134 mg/L, respectively. Only 2 % reduction in SS removal efficiency was observed at 4 hours compared with 6 hours of HRT.
Figure 4.4  Influent and effluent VSS concentrations and removal efficiencies of UAF2 at fluctuating HRT of 6 and 4 hours
4.3 COD removal

The tCOD and sCOD concentrations in the influent and effluent and their removal efficiencies in UAF1 and UAF2 under different HRTs were shown in Figures 4.5 and 4.6. Table 4.5 summarizes the average and range of influent and effluent tCOD and sCOD concentrations and their removal efficiencies.

The influent tCOD concentrations ranged from 343 to 803 mg/L, with an average value of 534 mg/L at 16 hours HRT, and effluent tCOD concentrations ranged from 34 to 282 mg/L, with an average value of 128 mg/L. The influent sCOD concentrations ranged from 45 to 211 mg/L, with an average value of 85 mg/L and the effluent sCOD concentrations ranged from 7 to 81 mg/L, with an average value of 49 mg/L. In addition, the removal efficiencies of tCOD and sCOD obtained were in the range from 58 to 92%, with an average of 77% and from 11 to 92%, with an average of 41%, respectively. As stated earlier in section 4.1, the removal efficiencies of tCOD and sCOD were below 50% during periodical biomass washouts at 16 hours HRT. Hence, the loss of biomass during wash out had reduced the COD removal efficiency.

At 12 hours HRT, the effluent tCOD and sCOD concentrations were higher than that at 16 hours HRT. This led to a 5% reduction in the tCOD removal efficiency and 20% reduction in sCOD removal efficiency. The reason for the lower removal efficiency could be due to higher organic loading rate at shorter HRT, which resulted in insufficient time to degrade the total and soluble COD.
Only 2% reduction in tCOD removal efficiency at 8 hours HRT was observed when compared with 12 hours HRT, while the sCOD removal efficiency experienced a 50% increase.
Table 4.5. Influent and Effluent tCOD and sCOD Concentrations and Removal Efficiencies

<table>
<thead>
<tr>
<th>Characteristics</th>
<th>HRT 16 hrs (UAF1)</th>
<th>HRT 12 hrs (UAF2)</th>
<th>HRT 8 hrs (UAF1)</th>
<th>HRT 6 hrs (UAF2)</th>
<th>HRT 4 hrs (UAF1)</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>Range</td>
<td>Ave</td>
<td>Range</td>
<td>Ave</td>
<td>Range</td>
</tr>
<tr>
<td>Influent tCOD (mg/L)</td>
<td>343 ~ 803</td>
<td>534</td>
<td>391 ~ 707</td>
<td>52</td>
<td>194 ~ 1134</td>
</tr>
<tr>
<td>Effluent tCOD (mg/L)</td>
<td>34 ~ 282</td>
<td>128</td>
<td>53 ~ 336</td>
<td>14</td>
<td>41 ~ 428</td>
</tr>
<tr>
<td>tCOD Removal (%)</td>
<td>58 ~ 92</td>
<td>77</td>
<td>29 ~ 88</td>
<td>72</td>
<td>44 ~ 93</td>
</tr>
<tr>
<td>Influent sCOD (mg/L)</td>
<td>45 ~ 211</td>
<td>85</td>
<td>37 ~ 211</td>
<td>90</td>
<td>32 ~ 191</td>
</tr>
<tr>
<td>Effluent sCOD (mg/L)</td>
<td>7 ~ 81</td>
<td>49</td>
<td>35 ~ 116</td>
<td>73</td>
<td>9 ~ 77</td>
</tr>
<tr>
<td>sCOD Removal (%)</td>
<td>11 ~ 92</td>
<td>41</td>
<td>4 ~ 62</td>
<td>21</td>
<td>21 ~ 89</td>
</tr>
</tbody>
</table>
Figure 4.5  Influent and effluent tCOD and sCOD concentrations and removal efficiencies of UAF1 at HRT 16, 8 and 4 hours
Figure 4.6 Influent and effluent tCOD and sCOD concentrations and removal efficiencies of UAF2 at HRT 12 and 6 hours.
The effluent tCOD and sCOD concentrations at 6 hours HRT were slightly higher than that at 16, 12 and 8 hours HRT. The reductions in removal efficiencies of tCOD and sCOD were 2% and 5%, respectively. A significant amount of reduction in tCOD removal efficiency (8%) was observed at 4 hours HRT compared with 6 hours HRT, while only 4% reduction in sCOD was noted.

Nebot et al. (1995) pointed out that the characterization of performance of the anaerobic filter versus organic load added is important. Several studies have been carried out to study the influence of the organic load shocks in the anaerobic filter. Consequently, Panswad and Komolmethee, (1997) mentioned that this shock can be produced by both the applied volumetric flow rate and the increase in the added organic load – these are called hydraulic shock or organic shock, respectively. The influent and effluent tCOD and sCOD concentrations and removal efficiencies of UAF2 at fluctuating HRT are shown in Figures 4.7 and 4.8, respectively. Table 4.6 shows the range and average of influent and effluent tCOD and sCOD concentrations and removal efficiencies at fluctuating loading rate.

The influent tCOD concentrations were in the range of 343 to 956 mg/L, with an average value of 579 mg/L and the effluent tCOD concentrations for 6 hours HRT and 4 hours HRT were ranged from 59 to 415 mg/L, with an average of 194 mg/L and 95 to 315 mg/L, with an average of 219 mg/L, respectively.
Table 4.6. Influent and Effluent tCOD and sCOD Concentrations and Removal Efficiencies of UAF2 at Fluctuating Loading Rate

<table>
<thead>
<tr>
<th>Characteristics</th>
<th>Range</th>
<th>Average</th>
<th>Range</th>
<th>Average</th>
</tr>
</thead>
<tbody>
<tr>
<td><strong>HRT = 6 hrs</strong></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Influent, tCOD (mg/L)</td>
<td>343 ~ 956</td>
<td>579</td>
<td>343 ~ 956</td>
<td>579</td>
</tr>
<tr>
<td>Effluent, tCOD (mg/L)</td>
<td>59 ~ 415</td>
<td>194</td>
<td>95 ~ 315</td>
<td>219</td>
</tr>
<tr>
<td>tCOD % reduction</td>
<td>20 ~ 89</td>
<td>65</td>
<td>30 ~ 86</td>
<td>59</td>
</tr>
<tr>
<td><strong>HRT = 4 hrs</strong></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Influent, sCOD (mg/L)</td>
<td>14 ~ 143</td>
<td>87</td>
<td>14 ~ 143</td>
<td>87</td>
</tr>
<tr>
<td>Effluent, sCOD (mg/L)</td>
<td>11 ~ 96</td>
<td>57</td>
<td>9 ~ 109</td>
<td>60</td>
</tr>
<tr>
<td>sCOD % reduction</td>
<td>15 ~ 81</td>
<td>33</td>
<td>10 ~ 77</td>
<td>30</td>
</tr>
</tbody>
</table>

The influent sCOD concentrations were in the range of 14 to 143 mg/L, with an average value of 87 mg/L and the effluent sCOD concentrations for 6 hours HRT and 4 hours HRT were ranged from 11 to 96 mg/L, with an average of 57 mg/L and 9 to 109 mg/L, with an average of 60 mg/L, respectively.

The removal efficiencies of tCOD for 6 hours HRT and 4 hours HRT were from 20 to 89%, with an average of 65% and from 30 to 86%, with an average of 59%, respectively. While the removal efficiencies of sCOD for 6 hours HRT and 4 hours HRT were ranged from 15 to 81%, with an average of 33% and from 10 to 77%, with an average of 30%, respectively. The observed COD removal results show that anaerobic filter performed similarly for 6 hours and 4 hours fluctuating HRT. This observation agreeded with Herbert et al. (1994) who reported that the percentage of COD removal from wastewater...
in an anaerobic reactor appeared to be mainly dependant on the volumetric COD loading rate, and relatively insensitive to the COD level in wastewater and the HRT.

**Figure 4.7**  Influent and effluent tCOD concentrations and removal efficiencies at fluctuating HRT of 6 and 4 hours
Figure 4.8  Influent and effluent sCOD concentrations and removal efficiencies at of UAF2 fluctuating HRT of 6 and 4 hours
4.4 BOD$_5$ Removal

The tBOD and sBOD concentrations of influent and effluent and their removal efficiencies of UAF1 and UAF2 under different HRTs are shown in Figures 4.9 and 4.10, respectively. The average and range of influent and effluent tBOD and sBOD concentrations and their removal efficiencies are summarized in Table 4.7.

At 16 hours HRT, the influent tBOD$_5$ concentrations were in the range of 138 to 436 mg/L, with an average value of 226 mg/L and the effluent tBOD$_5$ concentrations ranged from 10 to 85 mg/L, with an average value of 44 mg/L. The influent sBOD$_5$ concentrations ranged from 9 to 43 mg/L, with an average value of 24 mg/L and the effluent sBOD$_5$ concentrations ranged from 2 to 30 mg/L, with an average value of 13 mg/L. The removal efficiencies of tBOD$_5$ and sBOD$_5$ ranged from 59 to 94%, with an average of 79% and from 8 to 75%, with an average of 53%, respectively.

At 12 hours HRT, the removal efficiencies of tBOD$_5$ and sBOD$_5$ ranged from 48 to 91% with an average of 67% and from 25 to 41% with an average of 33%, respectively. There were 12% reduction in tBOD removal efficiency and 20% reduction in sBOD removal efficiency were observed at 12 hours HRT when compared with 16 hours HRT, which was comparable with Kobayashi et al. (1983) results (BOD removal 79%, COD removal 73%).
Figure 4.9. Influent and effluent \( tBOD_5 \) and \( sBOD_5 \) concentrations and removal efficiencies of UAF1 at HRT 16, 8 and 4 hours.
Figure 4.10 Influent and effluent tBOD$_5$ and sBOD$_5$ concentrations and removal efficiencies of UAF2 at HRT 12 and 6 hours
Table 4.7 Influent and Effluent tBOD and sBOD Concentrations and Removal Efficiencies

<table>
<thead>
<tr>
<th>Characteristics</th>
<th>HRT 16 hrs (UAF1)</th>
<th>HRT 12 hrs (UAF2)</th>
<th>HRT 8 hrs (UAF1)</th>
<th>HRT 6 hrs (UAF2)</th>
<th>HRT 4 hrs (UAF1)</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>Range</td>
<td>Ave</td>
<td>Range</td>
<td>Ave</td>
<td>Range</td>
</tr>
<tr>
<td>Influent tBOD₅ (mg/L)</td>
<td>138 ~ 436</td>
<td>22 6</td>
<td>172 ~ 245</td>
<td>210</td>
<td>141 ~ 346</td>
</tr>
<tr>
<td>Effluent tBOD₅ (mg/L)</td>
<td>10 ~ 85</td>
<td>44</td>
<td>16 ~ 127</td>
<td>71</td>
<td>29 ~ 142</td>
</tr>
<tr>
<td>tBOD₅ Removal (%)</td>
<td>59 ~ 94</td>
<td>79</td>
<td>48 ~ 91</td>
<td>67</td>
<td>54 ~ 89</td>
</tr>
<tr>
<td>Influent sBOD₅ (mg/L)</td>
<td>9 ~ 43</td>
<td>24</td>
<td>3 ~ 52</td>
<td>27</td>
<td>11 ~ 49</td>
</tr>
<tr>
<td>Effluent sBOD₅ (mg/L)</td>
<td>2 ~ 30</td>
<td>13</td>
<td>13 ~ 58</td>
<td>36</td>
<td>8 ~ 22</td>
</tr>
<tr>
<td>sBOD₅ Removal (%)</td>
<td>8 ~ 75</td>
<td>53</td>
<td>25 ~ 41</td>
<td>33</td>
<td>17 ~ 71</td>
</tr>
</tbody>
</table>
Likewise, while comparing removal efficiencies of tBOD and sBOD at 8 and 12 hours HRTs, there were significant reductions of 9 and 16 % in tBOD and sBOD at 8 hours HRT. There was no significant change in tBOD removal efficiency at 6 hours HRT compared with 8 hours HRT, however, a 7 % reduction in sBOD removal efficiency was observed.

Approximately, 5% reduction in removal efficiency of tBOD was observed at 4 hours HRT compared with 6 hours HRT. However, the sBOD removal efficiency experienced an increase of 5% at 4 hours HRT.

The influent and effluent tBOD and sBOD concentrations and removal efficiencies of UAF2 at fluctuating HRT are shown in Figures 4.11 and 4.12, respectively. Table 4.8 shows the range and average of influent and effluent tBOD and sBOD concentrations and removal efficiencies at fluctuating loading rate.
Figure 4.11  Influent and effluent tBOD\textsubscript{5} concentrations and removal efficiencies of UAF2 at fluctuating HRT 6 and 4 hours

Figure 4.12  Influent and effluent sBOD\textsubscript{5} concentrations and removal efficiencies of UAF2 at fluctuating HRT 6 and 4 hours
The influent $t\text{BOD}_5$ concentrations ranged from 111 to 311 mg/L, with an average value of 222 mg/L and the effluent $t\text{BOD}$ concentrations for 6 hours HRT and 4 hours HRT ranged from 35 to 148 mg/L, with an average of 64 mg/L and from 35 to 196 mg/L, with an average of 77 mg/L, respectively. The influent $s\text{BOD}_5$ concentrations ranged from 16 to 47 mg/L, with an average value of 27 mg/L and the effluent $s\text{BOD}_5$ concentrations for 6 hours HRT and 4 hours HRT ranged from 7 to 25 mg/L, with an average of 15 mg/L and from 7 to 34 mg/L, with an average of 18 mg/L, respectively.
The removal efficiencies of tBOD$_5$ for 6 hours HRT and 4 hours HRT ranged from 39 to 84% with an average of 71% and from 13 to 84% with an average of 65%, respectively. The removal efficiencies of sBOD$_5$ for 6 hours HRT and 4 hours HRT ranged from 18 to 75% with an average of 45% and from 12 to 60% with an average of 34%, respectively. The results show that the anaerobic filter performance at 6 hours and 4 hours fluctuating HRT in terms of tBOD$_5$ and COD removal did not differ significantly. It could be concluded that anaerobic filter performed similarly at 6 hours and 4 hours fluctuating HRT in terms of the solids, COD and BOD$_5$ removal. In addition, this study demonstrated that the anaerobic filter was able to withstand shock loadings and perform well with the help of the attached biofilm.
4.5 Biogas Composition

Biogas composition of UAF2 at HRT of 12 and 6 hours are shown in Figure 4.13, while Figure 4.14 illustrates the biogas composition of UAF1 at HRTs of 16, 8 and 4 hours.

During the start up of UAF1 at HRT of 16 hours (221 days operation, from Day 20 to Day 241), methane composition in the biogas increased gradually from 2.41 to 50.50% until Day 150. Thereafter, the methane composition was relatively stable and maintained at around 50%. While at HRT of 12 hours (operated in UAF2) (75 days operation, from Day 1 to Day 75), the methane composition showed gradual increase from 4.13 to 29.71%, during the 75 days of operation. Further reducing the HRT to 8 hours (in UAF1) demonstrated that higher methane composition was obtained (ranged from 65.96 to 77.98% and stabilized at 76.5 %). This was due to the higher organic loading rate at shorter HRT. However, the maximum methane composition obtained at 6 hours HRT was approximately 13% more than that achieved at 8 hours HRT (the methane composition at 6 hours HRT increased gradually from 4.45 to 63.89% and stabilized at around 63% from Day 240 onwards). As UAF2 was reseeded and methanogens are slow growing bacteria, lower methane gas production of 4.45% was observed during the initial period. When operated at 4 hours HRT, the measured methane composition ranged from 77.22 to 77.32% and stabilized at 77.23 %. This value was slightly higher than that at HRT of 8 hours, as organic loading was higher while reducing the HRT.
Figure 4.13. Biogas composition of UAF2 at HRT of 12 and 6 hours.
**Figure 4.14.** Biogas composition of UAF1 at HRT of 16, 8 and 4 hours.
4.6 Biogas Production

Biogas production of UAF1 at HRTs of 16, 8 and 4 hours and UAF2 at HRTs of 12 and 6 hours are shown in Figures 4.15 and 4.16, respectively.

During the UAF1 start up period at 16 hours HRT, biogas production increased gradually from 0.09 to 0.97 L/d. On Day 139, biogas leakage near effluent port was noted and was later replaced with a T-junction setup. After the retrofitting, an increase in the biogas production volume was observed. From Day 150, the biogas production was rather stable at around 0.75 L/d.

Similarly, the biogas production also showed gradual increase during the UAF2 start up period at 12 hours HRT within the 75 days operation. Higher biogas production was observed at 8 hours HRT in UAF1 due to the higher organic loading rate. The biogas production ranged from 0.97 to 2.62 L/d and stabilized at around 2 L/d. Hence, shortening the HRT by half had contributed to an increase in biogas production by almost 2.6 times. However, due to reseeding of UAF2 at 6 hours HRT, the biogas production increased gradually from 0.88 to only up to 1.46 L/d and eventually only stabilized at around 1.42 L/d from day 115. Further shortening the HRT to 4 hours gave rise to biogas production ranged from 2.12 to 2.63 L/d and stabilized around 2.19 L/d. The biogas production at 4 hours HRT was only 9.5% more than that at 8 hours HRT.
Figure 4.15. Biogas production of UAF1 at HRT of 16, 8 and 4 hours.
Figure 4.16. Biogas production of UAF2 at HRT of 12 and 6 hours.
4.7. Suspended Biomass (MLSS and MLVSS)

The suspended biomass concentration of anaerobic filter was measured in terms of MLSS and MLVSS. The suspended biomass sample was collected from the sampling port located 0.37 m from the top of the reactor. The suspended biomass concentrations of UAF2 at HRT of 12 and 6 hours and UAF1 at HRT of 16, 8 and 4 hours are shown in Figures 4.17 and 4.18, respectively.

During the UAF1 start up at HRT of 16 hours, the MLSS and MLVSS concentrations ranged from 2411 to 8517 mg/L with an average of 5543 mg/L and from 2000 to 6167 mg/L with an average of 3794 mg/L, respectively. The MLSS and MLVSS concentrations during the UAF2 start up at HRT of 12 hours were significantly higher which ranged from 5360 to 13073 mg/L with an average value of 7841 mg/L and from 3673 to 8209 mg/L with an average value of 5340 mg/L, respectively. Consequently, during 8 hrs of HRT, the MLSS and MLVSS concentrations were observed to range from 3567 to 13,856 mg/L and from 2600 to 9856 mg/L, respectively. And the average MLSS and MLVSS concentrations were 6374 and 5022 mg/L at 8 hours of HRT. While comparing MLSS and MLVSS concentrations of 16 and 8 hours HRTs, an increase in concentrations of 13% and 24%, respectively, was observed due to higher organic loading at 8 hours HRT than 16 hours. The reasons for this could be higher hydraulic loading rate would have sloughed off the attached biofilm and higher organic loading would have increased the suspended biomass concentration in UAF1.
Moreover, during 6 hrs of HRT, the MLSS and MLVSS ranged from 4067 to 17,722 mg/L and from 3,467 to 12,644 mg/L, respectively. And the average MLSS and MLVSS concentrations were 10,110 and 8058 mg/L at 6 hrs of HRT. Finally, during 4 hrs of HRT, the MLSS and MLVSS ranged from 4267 to 7656 mg/L and from 2711 to 7111 mg/L, respectively. And the average MLSS and MLVSS concentrations were 6265 and 4758 mg/L at 4 hrs of HRT.

The general trend shows that reducing the HRT resulted in increased MLSS and MLVSS in the anaerobic filter. Shortening the HRT had increased the loading rate which in turn provided more organics and nutrient that stimulated higher growth rate.

Figure 4.17. MLSS and MLVSS concentrations of the suspended biomass of UAF2 at HRT of 12 and 6 hours.
Figure 4.18. MLSS and MLVSS concentrations of the suspended biomass of UAF1 at HRT of 16, 8 and 4 hours.
4.8 Attached Biomass (MLSS and MLVSS)

Henze and Harremoes (1982) indicated that there is a wide variation in the attached biomass content in the upflow anaerobic filters. In addition, they observed that 80% of the biomass in an anaerobic filter was attached and the rest was suspended.

Miyahara et al. (1995) found that the number of suspended acidogenic bacteria in an AF reactor was higher than that of bacteria attached to the media. On the other hand, the number of attached methanogenic bacteria was higher than the number of suspended methanogenic bacteria.

The attached biomass concentration in the UAF2 reactor was measured at Day 200 by taking two samples of media. After washing with tap water, the biomass concentration in the washing water from each sample was measured. Based on these data, the MLSS and MLVSS of attached biomass concentrations in the UAF2 reactor were estimated as 18,000 mg/L and 14,200 mg/L at 12 hours HRT. This observation agreed with Emitwalli et al. (2002b) who revealed that MLVSS concentration of 15,100 mg/L. Subsequently, the attached biomass contains MLSS of 12,654 mg/L and MLVSS of 10,743 mg/L at 6 hrs of HRT. While comparing 12 and 6 hrs of HRT, it was observed that there was a reduction in attached biomass concentration. Hydraulic effect of higher hydraulic load sloughed off of the attached biomass.
4.9 Mass balance (carbon balance) of UAF1 at HRT 8 hours

The following experimental data of steady conditions are used for the calculation and estimation.

Flowrate \( Q = 61.5 \) L/d

Avg. \( \text{CH}_4 \) % = 76.5%

Avg. biogas production = 2.0 L/d

Avg. \( \text{tCOD}_{\text{inf}} \) = 579 mg/L

Avg. \( \text{tCOD}_{\text{eff}} \) = 168 mg/L

Calculation of \( \text{CH}_4 \) dissolved in the liquid (effluent)

\( \text{CH}_4 \) solubility at 1 atm \( \sim \) 20 mg/L

\( \text{CH}_4 \) partial pressure = 0.765 atm (as UASB avg. \( \text{CH}_4 \) % = 76.5%)

Dissolved \( \text{CH}_4 \) = 0.765 \times 20 \text{ mg/L} = 15.3 \text{ mg/L} = 0.9563 \text{ m mol/L}

Flowrate \( Q = 61.5 \) L/d

Dissolved \( \text{CH}_4 \) in the effluent (loss) = 61.5 L/d \times 0.9563 \text{ m mol/L} = 58.81 \text{ m mol/d} = 0.05881 \text{ mol/d}

Avg. biogas production = 2.0 L/d (biogas collected)

\( \text{CH}_4 \) production = 2.0 L/d \times 76.5\% = 1.53 L/d (\text{CH}_4 \) collected at 1 atm & \( 30^\circ\text{C} \)
Based on PV = nRT

\[ P_1 V_1 / T_1 = P_0 V_0 / T_0 \]

At standard condition: \( P_0 = 1 \text{ atm}, \ T_0 = 0^\circ\text{C} = 273 \text{ K}, \ 1 \text{ mol gas} = 22.4 \text{ L} \)

\( P_1 = 1 \text{ atm}, \ V_1 = 1.53 \text{ L/d}, \ T_1 = 30^\circ\text{C} = 303 \text{ K} \)

\( V_0 = 1.3785 \text{ L/d (at standard condition)} = 0.06154 \text{ mol/d} \)

The amount of CH\(_4\) collected \( n = 0.06154 \text{ mol/d} \)

Loss rate of CH\(_4\) dissolved in the liquid (effluent)

\[ = \frac{(0.05881 \text{ mol/d})}{((0.06154 \text{ mol/d}) + (0.05881 \text{ mol/d}))} \]

\[ = 48.87\% \]

Mass balance calculation

Total COD\(_{\text{in}}\) = \( t\text{COD}_{\text{int}} \times \text{Flowrate Q} = 579 \text{ mg/L} \times 61.5 \text{ L/d} = 35608.5 \text{ mg/d} = 35.61 \text{ g/d} \)

There is no sludge wasting in the AnF operation, so waste sludge-COD = 0.

The loss due to sampling is ignored.

At steady conditions, assume no reactor biomass change.

If the COD loss as CO\(_2\) is not considered, then

Total COD\(_{\text{out}}\) = \( (t\text{COD}_{\text{eff}} \times \text{Flowrate Q}) + \text{CH}_4\)-COD

\( t\text{COD}_{\text{eff}} \times \text{Flowrate Q} = 168 \text{ mg/L} \times 61.5 \text{ L/d} = 10332 \text{ mg/d} = 10.332 \text{ g/d} \)
Based on \( \text{CH}_4 + 2\text{O}_2 \rightarrow \text{CO}_2 + 2\text{H}_2\text{O} \)

\[
\text{CH}_4 \text{ collected } = 1.3785 \text{ L/d } = 0.06154 \text{ mol/d} \\
\text{CH}_4\text{-COD collected } = 2 \times 0.06154 \text{ mol/d } = 0.1231 \text{ mol/d } = (0.1231 \times 32) \text{ g/d } = 3.939 \text{ g/d}
\]

Because the loss rate of \( \text{CH}_4 \) dissolved in the liquid (effluent) = 48.87\%, then

Total \( \text{CH}_4\text{-COD} \) = 7.704 g/d

Total \( \text{COD}_{out} \) = 10.332 g/d + 7.704 g/d = 18.036 g/d

Recovery rate = \((18.036 \text{ g/d}) / (35.61 \text{ g/d}) = 51\%\)

### 4.10 Operational Problems:

Gas bubbles entrapped in filters may cause channeling and short-circuiting in the reactor (Henz and Harremoes, 1982). In full scale studies, filter clogging has been reported after 18 months of continuous operation (Subbiah, 1997). Kobayashi et al (1983) operated two larger scale filters 0.6 m in diameter by 2.6m high) and experienced no clogging problems in three years of operation. Similarly, operational problems like channeling, short-circuiting and clogging were not observed during the entire period of this research study. However, additional work is still required to evaluate clogging problems.
5 Conclusion & Recommendations

5.1 Conclusion

It has been shown that an anaerobic filter operating at high and low loading rate can successfully treat domestic wastewater at temperature of 30°C. The results show that the anaerobic filter performance at different HRTs in terms of tBOD$_5$ and COD removal did not differ significantly. It could be concluded that anaerobic filter performed similarly at 6 hours and 4 hours fluctuating HRT in terms of the solids, COD and BOD$_5$ removal. In addition, this study demonstrated that the anaerobic filter was able to withstand shock loadings and perform well with the help of the attached biofilm. Moreover, effluent quality did not achieve secondary effluent quality standards but came very close. Anaerobic process could also be followed by aerobic processes for effluent polishing to utilize the benefits of both processes. Series reactors of anaerobic-aerobic processes have been feasible for treating municipal wastewaters in warmer climates resulting in lower energy requirements and less sludge production. It could be concluded that the anaerobic filter is a promising candidate for treatment of low strength wastewaters and that post treatment for sulfides and ammonia may be necessary.
5.2 Recommendations

- Start-up of anaerobic reactors is more time consuming and is subjected to disturbances more than that of aerobic reactors. Many researchers have reported long start-up periods of 2-3 months to 1 year (or even more) for the anaerobic reactors. The start-up of the anaerobic process is still considered as a major area of research. Considerable efforts have been made to study the granulation process but the mechanism involved in the formation of granulation sludge is still unknown. Therefore, it is recommended that further research have to be done on start-up of anaerobic process and granule formation of sludge.

- Study on UAF with separation of phases can be carried out. Separation of phases enables selection of optimal conditions for both processes – acidogenesis/fermentation and acetogenesis/methanogenesis. As fermentation (pH 5.2 – 5.9) proceeds at a much greater speed than acetogenesis/methanogenesis (pH 7.3 – 7.6), the former is carried out in acidic conditions in a separate reactor. So, it is recommended that further study have to be done on UAF with phase separation.

- It is well known that the operational state of high-rate anaerobic system can change within a few hours in response to a process disturbance. On the other hand, because of the special hydrodynamics in the reactors, the response of the
system output is deferred corresponding to an input variation. In order to provide prompt corrective response, which is required in case of organic overload shocks or toxic events, techniques that provide responses are needed to avoid the significant process deterioration and failure. In order to be effective for on-line control, a model can be developed to make predictions of the system response to the variations of organic, hydraulic and alkalinity loading rates. Based on the observed results a Monod-type of kinetic model can be developed to predict the performance of the reactor and methane content in produced biogas.

- The use of an adequate recycling ratio is a key factor for the optimization of the anaerobic treatment of these wastewaters, since it increases mass transfer rates, minimizes dead zones and allows development of more equilibrated consortia of bacteria, belonging to different trophic groups, along the reactor. It is recommended that further study have to be done on UAF with phase separation. Therefore, it is recommended that further research have to be done on UAF with recycling ratio.

6. REFERENCES:


APPENDIX - 1

1. EPS EXTRACTION METHOD

Principle

EPS was separated from the microorganism cell wall by using cation resin exchange. Cation exchange resin will remove cations from the sludge matrix leading to a break up of the flocs and a subsequent release of EPS.

Reagents

1. Phosphate buffer – 9mM NaCl, 1mM KCl, 2mM Na\textsubscript{3}PO\textsubscript{4} and 4mM NaH\textsubscript{2}PO\textsubscript{4} at pH 7 (526 mg/L NaCl, 74.56 mg/L KCl, 328 mg/L Na\textsubscript{3}PO\textsubscript{4} and 480 mg/L NaH\textsubscript{2}PO\textsubscript{4} in 1L DI water)

2. Cation exchange resin (CER) – Dowex Marathon C

Procedure

1) Wash the CER in phosphate buffer (1 kg CER in 2L phosphate buffer) before use. (Stir for 1 hour)
2) Take 200 mL of sludge sample, centrifuge at 4° C and 9000 RPM for 10 min.
3) Decant the supernatant.
4) Resuspend the sludge pellet to the original volume using phosphate buffer.
5) Transfer the suspension to an open-mouth closed container.
6) Add 70g CER /g VSS in a closed container.
7) Stir the suspension at 600 RPM and 4° C for 1.5 hr.
8) Centrifuge the suspension at 9000 RPM for 10 mins to separate the CER and biomass.
9) Collect the supernatant for subsequent analysis of EPS.
2. Measuring SS and VSS

1. Weigh the crucible with filter paper as W1.
2. Filter 100 ml of sewage sample using the filter paper.
3. Place the crucible with filter paper into the oven at 105\(^0\)C for 1 hour.
4. Take it out from the oven and cool it in a dessicator for an hour.
5. Measure it as W2.
6. Place it in a furnace at 550\(^0\)C for 20 minutes.
7. Cool it down in a dessicator for 1 hour.
8. Measure it as W3.

<table>
<thead>
<tr>
<th>Crucible 1 (g)</th>
<th>Crucible 2 (g)</th>
</tr>
</thead>
<tbody>
<tr>
<td>W1</td>
<td>W2</td>
</tr>
<tr>
<td>29.2306</td>
<td>29.2376</td>
</tr>
</tbody>
</table>

Formula:

\[ SS = W2 - W1 \]
\[ VSS = W2 - W3 \]

\[ SS_1 = 70 \text{ mg/l} \]
\[ SS_2 = 73 \text{ mg/l} \]

\[ VSS_1 = 64 \text{ mg/l} \]
\[ VSS_2 = 70 \text{ mg/l} \]

\[ \text{Ave. SS} = 71.5 \text{ mg/l} \]
\[ \text{Ave. VSS} = 67 \text{ mg/l} \]